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Battelle Columbus Labs.

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April 1976

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16. ABSTRACT The report gives results of an analysis of the constraints and environmental impact of converting selected industries to low- and intermediate-energy gas from coal. Two target industries, the secondary steel industry and the petroleum refining industry, were selected for analysis. A hypothetical model plant was developed for each target industry and characterized as to layout, energy use, combustion process characteristics, and relation to the respective target industry as a whole. A gasifier and gas-cleaning system combination was selected for each model plant and sized to provide sufficient low- or intermediate-energy gas to replace the model plant's requirement for natural gas and oil. Material and energy balances were made for each model plant, and the constraints involved in process modification, along with the potential environmental impact, were evaluated.

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Coal Gasification		Stationary Sources	13H
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SUMMARY

This study involved an analysis of the constraints and environmental impact of converting selected industries to low- and intermediate-energy gas from coal. Two target industries (the secondary steel industry and petroleum refining industry) were selected for analysis. A hypothetical model plant was developed for each target industry and characterized as to layout, energy use, combustion process characteristics, and relation to the respective target industry as a whole. A gasifier and gas-cleaning system combination was selected for each model plant and sized to provide sufficient low- or intermediate-energy gas to replace the model plant's requirement for natural gas and oil. Material and energy balances were done for each model plant, and the constraints involved in process modification, along with the potential environmental impact, were evaluated.

The model steel plant had a capacity of 996,000 metric tons (1,100,000 tons) of molten steel per year and an average demand for natural gas and oil of 21.1 x 10^6 MJ/day (20 x 10^9 Btu/day). A Kopper-Totzek gasification plant with four 2-head gasifier units combined with a MDEA (methyldiethanolamine) gas-cleaning system was selected to provide intermediate-energy gas (HHV 11.27 MJ/NM^3; 285 Btu/scf) for the plant.

The model refinery plant had a capacity of 3,972,500 liter/day (25,000 barrel/day) of crude oil and an average energy demand for natural gas and oil of 4.13 x 10^6 MJ/day (3.92 x 10^9 Btu/day). A Wellman-Galusha gasification plant with three 10-foot diameter gasifiers combined with a Stretford gas-cleaning system was selected to provide low-energy gas (HHV 6.62 MJ/Nm^3; 1062 Btu/scf), yielding a product gas with a high-heating value varying from 9.26 MJ/Nm^3 (235 Btu/scf) during the winter to 10.8 MJ/Nm^3 (274 Btu/scf) during the summer.

It was concluded that there were no major technological constraints in converting the model secondary steel plant to intermediate-energy gas or the model refinery to low-energy gas (when mixed with refinery gas). In both cases, however, most burners and gas distribution networks would have to be replaced. Based on current data and knowledge, there appeared to be no insurmountable environmental problems in retrofitting either model plant to gas from coal providing appropriate commercially available control equipment is employed. However, studies will be necessary to acquire new data to define the real environmental impact which might require more or less control for new future sources.

LIST OF ABBREVIATIONS

scf - Standard cubic foot as measured at 60 F and 30 inches of mercury (American Gas Association standard temperature and pressure)

psia - Pounds pressure per square inch absolute

psig - Pounds pressure per square inch gage

osi - Ounces pressure per square inch gage

Kcal - Kilocalorie

J - Joule

MJ - Mega joule (10^6 joule)

Kg - Kilogram

lb - Pounds mass

Btu - British Thermal Unit

m - Meter

mm - Millimeter

Nm^3 - Normal cubic meter

C - Degrees Celsius

F - Degrees Fahrenheit

N - Newton

ppm - Parts per million by volume (at 0 °C and 760 mm Hg)

HHV - High heating value of fuel including the latent heat of vaporization of water formed during combustion

LHV - Low heating value of fuel not including the latent heat of vaporization of water formed during combustion

LIST OF CONVERSION FACTORS

Btu (at 60 F) x 0.252×10^3 = calorie (cal)

Btu (at 60 F) x 1.055×10^3 = Joule (J)

feet x 0.3048 = meter (m)

degrees Fahrenheit (F) -32 x 0.555 = degrees Celsius (C)

standard cubic foot (scf) (at 60 F and 30 in. Hg) x .0268 = normal cubic meter (Nm^3) (at 0 ^{o}C and 760 mm Hg)

Btu/scf x 0.0394 = Mega Joule/Nm^3 (MJ/Nm^3)

pound mass (lb) x 0.453 = Kilogram (Kg)

lb/10^6 Btu x 11.798 = Kg/10^6 Kcal

U.S. ton (2000 lb) x 0.906 = metric ton (1000 Kg)

pound force per square inch (psi) x 6.89×10^3 = Pascal (Pa) = Newton/m^2 (N/m^2)

psi x 7.03×10^2 = Kg force/m^2

ounces per square inch (osi) x 0.431×10^3 = Pa

grains x 6.5×10^{-5} = Kg

gallon (U.S.) x 3.78 = liter

barrel (42 gallon) x 158.97 = liter

acre x 4050 = m^2

INTRODUCTION

In the past several decades, U.S. industry has become increasingly reliant on natural gas and petroleum as sources of energy. Increased industry use of natural gas and oil as prime fuel resulted from: (1) the availability of these fuels at low cost; (2) the low purchase, operating, and maintenance cost of gas- and oil-fired equipment; (3) the relatively dependable supply of gas and oil such that large fuel storage facilities were not needed; and (4) the environmental acceptability of these fuels. As can be seen from Table 1, industry accounted for about 45 percent of the natural gas, 17 percent of the petroleum, and 32 percent of the total energy consumed in the United States in 1972.

TABLE 1. U.S. TOTAL AND INDUSTRIAL ENERGY CONSUMPTION IN 1972(1)

	Energy Consumed, 10^{12} MJ (10^{15} Btu)		
	Natural Gas	Petroleum	Total
Industry	11.0 (10.4)	5.8 (5.5)	24.1 (22.9)
Total U.S.	24.4 (23.1)	3.5 (33.0)	76.0 (72.1)

In recent years, industry has been faced with acute shortages of natural gas as a source of energy. As a result, many industries have become increasingly reliant on oil and propane, the easiest and most immediately available substitutes for natural gas. Increased demand for these alternative fuels, however, along with the limited domestic supply and foreign politics, have caused their cost to increase dramatically and their availabitliy to be uncertain.

Projections as to the future supply and availability of natural gas vary. A recent projection is shown in Figure 1. The general conclusions are, however, that supply will continue to be limited and that, either through allocation or increased price, natural gas will be increasingly unavailable and unattractive as an industrial fuel. Oil is currently available to industry as an alternative fuel to natural gas; however, the availability of this fuel is largely a function of extensive imports. In the future it is probable that there will be increasing private and Government pressure to reduce these imports. Domestic supply of oil, on the other hand, is limited, with little hope for significant increase supply in the future, as shown in Figure 2.

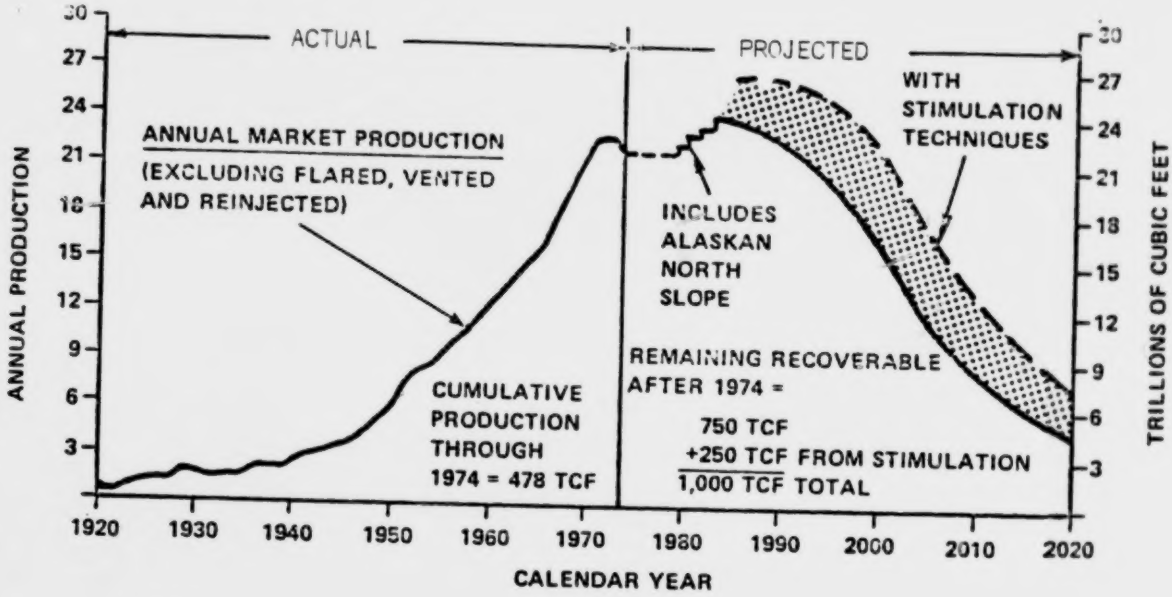

FIGURE 1. PROJECTED DOMESTIC NATURAL GAS PRODUCTION(2)

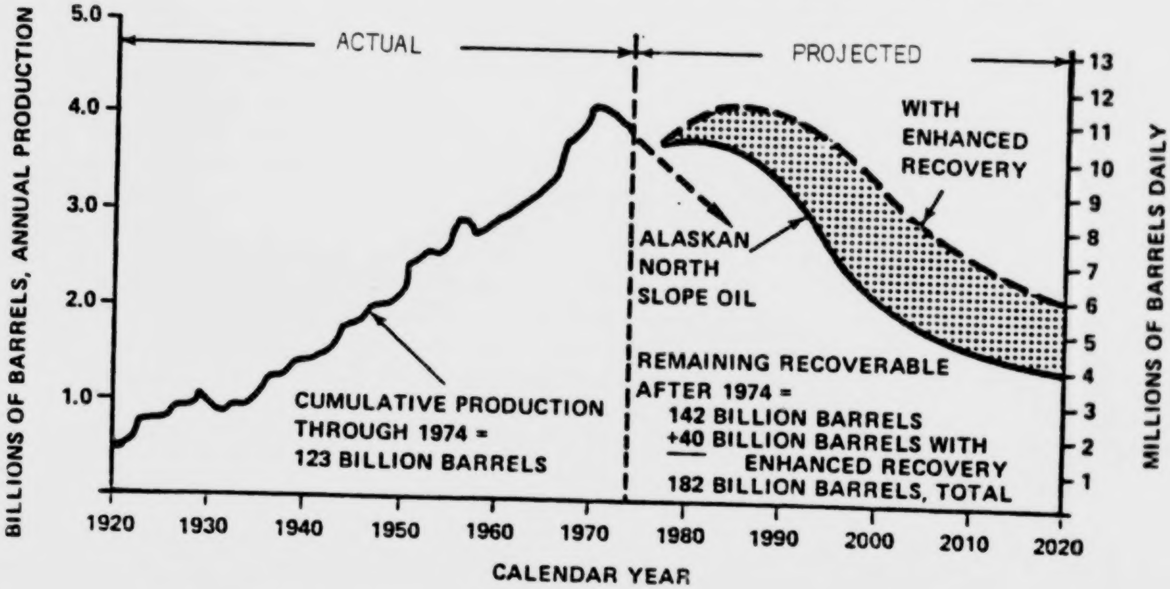
IN THIS FIGURE, DOMESTIC OIL INCLUDES CRUDE AND NATURAL GAS LIQUIDS

FIGURE 2. PROJECTED DOMESTIC OIL PRODUCTION(2)

A common recommendation is for industry to convert fuel-using processes to America's most abundant fossil-fuel resource, coal. As can be seen in Table 2, coal resources dwarf those of natural gas and oil in domestic supply.

TABLE 2. DOMESTIC FOSSIL-FUEL RESERVES(2)

Resource	Recoverable Reserves 10^{12} MJ	(10^{15} Btu)
Coal	12,700	(12,000)
Natural Gas	818	(775)
Petroleum	840	(800)

But, converting to direct firing of coal in boilers or processes may require extensive alteration and the addition of pollution control equipment. Also, many types of industrial processes are not readily adaptable to direct firing of coal.

An alternative to direct use of coal is to convert the coal to a clean liquid or gaseous fuel prior to firing it. The technology for doing this is not new. Reportedly there were over 11,000 coal-gasification plants in the United States in the 1920's making a low-heating value fuel gas for a wide variety of industrial uses(3). Only a few of these plants remain in use today.

A variety of types of gasification processes were used for making gas from coal in the past and many of these are still commercially available. Basically the gas from these processes could be categorized by heat content as either low-energy gas, having a heating value of about 4.7 to 7.9 MJ/Nm^3 (120 to 200 Btu/scf), or intermediate-energy gas, having a heating value of about 9.85 to 13.8 MJ/Nm^3 (250 to 350 Btu/scf). Low-energy gas is made by reacting air and steam with coal in a partial combustion process yielding a gas primarily composed of CO_2, CO, H_2, and approximately 50 percent N_2. When oxygen is substituted for the air, the N_2 is reduced to about 1 to 2 percent and the heating value of the fuel gas is doubled. The intermediate-energy gas can be used as a fuel or as the synthesis gas for production of methane or higher hydrocarbon products. Most gasifiers can produce either low- or intermediate-energy gas. However, some slagging units are limited primarily to oxygen-blown operation due to the higher temperatures required to maintain slagging conditions.

The primary reason this technology was practically abandoned in this country was the availability of natural gas and oil. These high-grade fuels were cleaner, easier to use, less hazardous to handle (gas from coal contains toxic CO), and above all, less expensive. Their use has prompted the development of more sophisticated burners and furnaces with precise heat release characteristics and better control systems which allow safer, more efficient operation with a minimum of operator attention. Modern glass melting operations, for instance, require about half the melter area for a given production rate as melters did 30 years ago--when firing with gas from coal was common. This increase in productivity is a result of many improvements, but the availability of natural gas and oil was an important factor.

Converting industries from natural gas to gas from coal involves many considerations, including fuel-gas production, utilization of the fuel gas, and the environmental impact of gasifiers. Two of these considerations relate to coal gasification for any application (power plant, industry, etc.), however, the utilization aspect is much more critical for retrofitting industrial applications where a wide variety of process characteristics must be considered. For instance, low- and intermediate-energy gas has a heating value of from about one-sixth to one-third that of natural gas. In addition, other combustion properties such as flame temperature, burning velocity, and radiation characteristics will also be different than those of natural gas.

This report describes the potential for conversion of two model industrial plants from a primary dependence on natural gas and oil to the production and utilization of fuel gas produced on-site from coal. Gasification, utilization, environmental, and economic aspects are discussed with the primary purpose being to assess the potential environmental problems and to aid in determining the priority that limited environmental resources should have in these areas.

OBJECTIVE

The objective of this study was to quantitatively evaluate the potential environmental impact of retrofitting selected American industries from the use of natural gas and oil to the use of low- and intermediate-energy fuel gas produced from coal and to quantify the major constraints and problems that would be encountered in such a retrofit.

APPROACH

Two industries were selected as target industries based on an analysis of their amenability to low- or inetermediate-energy gas. Hypothetical model plants were then developed based on the characteristics of typical plants within each of the target industries. Each of the model plants was characterized as to its existing fuel-use and fuel-distribution patterns, existing pollution abatement systems, and location relative to coal supply and reserves.

Subsequently, for each model plant, a gasification system (including gasifier and gas-cleaning processes) was selected on the basis of commercial availability, operating limitations, unit size, and the potential for its integration into the model plant, including the potential for use of byproducts or wastes. The impact of installing the gasification plant and converting the plant to distribution and utilization of the fuel gas were determined, and an assessment was made of the environmental impact of converting the plant to coal gasification.

CONCLUSIONS

It was concluded that there were no major technological constraints in converting the model secondary steel plant to intermediate-energy gas or the model refinery to low-energy gas (when mixed with refinery gas). In both cases, however, most burners and gas distribution networks would have to be replaced. Based on current data and knowledge, there appeared to be no insurmountable environmental problems in retrofitting either model plant to gas from coal providing appropriate commercially available control equipment is employed. However, studies will be necessary to acquire new data to define the real environmental impact which might require more or less controls for new future sources.

Sulfur and nitrogen compounds in the fuel gas (primarily H_2S and NH_3) represent the primary source of atmospheric emissions from combustion of fuel gas. These constituents are also potentially corrosive, especially if any water vapor is present. Although the corrosivity of these constituents has not been accurately defined, it appears that their concentration in the fuel gas should be reduced below that which would be considered environmentally acceptable to minimize corrosion in the intricate and extensive gas distribution systems found in most industrial plants.

The major environmental hazard involves the gasification/gas cleanup plant itself. Many of the potential atmospheric emissions in the fuel gas become potential liquid and solid effluents after they are removed from the gas. A variety of processes will be required for treating the various liquid streams used in cleaning and cooling the fuel gas. In many industries water treatment systems are already in use similar to those that would be required for a gasification plant. In the refinery industry, for example, water treatment processes are commonly used for removing oils, phenols, ammonia, and H_2S; however, in almost all cases these treatment systems would have to be enlarged or additional processes added if a gasification plant was installed. The expertise and technology for treating these various streams in an environmentally acceptable manner appear to be avialable.

The major factor in determining whether an industry would convert to low- or intermediate-energy gas is economics. Even a small gasification plant would involve many processes, most of which involve either cleaning the gas or treating various effluent streams associated with gas cleaning. These various cleaning systems also constitute the major cost in a coal gasification plant. Modifications required in furnaces are primarily determined by the heating value of the gas with high heating value gases requiring fewer modifications. The requirements for a complex gasification plant and attractiveness of a high grade fuel gas (such as intermediate-energy gas from an oxygen-blown gasifier) tend to favor large-scale industrial applications. Thus, gasification is most attractive for large individual industrial plants or groups of smaller plants in an industrial park.

I. TARGET INDUSTRY SELECTION

An important part of this study of industrial energy/fuel problems was the selection of the target industries for analysis. To determine those industries to be included, various four-digit SIC code industrial groups were evaluated as to their applicability to low- and intermediate-energy gas according to the following industry selection criteria.

(1) Consumption of Natural Gas and Fuel Oil. The selected industries should be major consumers of natural gas and fuel oil on a national basis. Selection of such industries would insure that, should they convert to low- or intermediate-energy gas from coal, there would be a significant impact on releasing natural gas and oil for other uses.

(2) Amenability of the Processes to Low- and Intermediate Energy Gas. The industries selected should have processes and plant sizes that would make them appear amenable to conversion to low- or intermediate-energy gas. Results of a previous survey study on converting industrial processes to low-energy gas were used in evaluating industries relative to these criteria(4).

(3) Location Relative to Available Coal Supplies. Any industry selected should be generically located near coal supplies. Because this study involved only on-site gasification, coal would have to be shipped to the plant.

(4) Dependence on Natural Gas and Oil. Any industry selected should have a high degree of dependence on a source of clean gaseous or liquid fuels. Those industries that could more easily convert to direct use of coal would be considered less urgent for study than those that could not.

(5) Potential for Long-Term Utilization of Low- and Intermediate-Energy Gas. The cost of energy is important to all industries, however, some industries are more energy intensive than others and as a result, are more sensitive to energy costs. These energy-intensive industries were felt to have more incentive to make the long-term commitment necessary in installing a coal gasification facility. Less energy-intensive industries, on the other hand, would be more likely to pay high prices for premium fuels (remaining natural gas and oil, electricity, or future high-grade fuels from coal) to minimize the amount of modification necessary in their processing operations.

Evaluation of Candidate Industries

A list of the major energy consuming industries (by four-digit SIC code) was compiled for evaluating candidate industries. Table 3 lists 17 industries including the 10 major consumers of natural gas, distillate and residual fuel oil, and coal in 1971. After initial screening of this list the following industries were selected for further study:

Petroleum Refining	SIC 2911
Blast Furnaces and Steel	SIC 3312
Industrial Organic and Inorganic Chemicals	SIC 2818, 3819
Hydraulic Cement	SIC 3241
Paper and Paperboard Mills	SIC 2621, 2631
Primary Aluminum	SIC 3334
Glass Containers	SIC 3221.

These seven industry groups were then ranked in order for each of the five criteria. The results of this ranking are shown in Table 4. The ranking is somewhat arbitrary and based on views gained from a variety of sources of information resulting from this study and the previous study on converting industrial processes to low- and intermediate-energy gas(4).

According to this ranking, two industry groups, petroleum refining (SIC 2911) and blast furnaces and steel mills (SIC 3312), appeared highly attractive for selection as target industries. Discussions were held with representatives of both of these industries and, subsequently, these industries were selected for detailed study.

The analysis used in evaluating the two selected target industries under the five criteria was as follows.

TABLE 3. ENERGY USE BY THE 10 MAJOR INDUSTRIAL CONSUMERS OF GAS, OIL, AND COAL BY 4-DIGIT SIC CODE FOR 1971(1)

SIC		Ranking			Fuel Use, 10^9 MJ/year (10^{12} Btu/year)(a)				Total Energy Use(c)
		Gas	Oil	Coal	Gas	Oil	Coal	Total(b)	
3312	Blast Furnaces and Steel	2	1	3	689 (653)	176 (167)	140 (133)	1005 (953)	1323 (1254)
2818	Industrial Organic Chemicals (not elsewhere classified)	3	10	4	638 (605)	27 (26)	139 (132)	805 (763)	958 (908)
3241	Hydraulic Cement	5	5	1	219 (208)	43 (41)	190 (180)	452 (429)	452 (459)
2621	Paper Mills	6	2	2	212 (201)	170 (161)	157 (149)	569 (511)	555 (526)
2631	Paperboard Mills	7	3	5	191 (181)	166 (157)	86 (82)	443 (420)	460 (436)
2911	Petroleum Refining	1	4	-	1405 (1332)	72 (68)	9.5 (9)	1486 (1409)	1509 (1440)
2819	Industrial Inorganic Chemicals (not elsewhere classified)	4	-	-	396 (375)	24 (23)	37 (35)	457 (433)	478 (453)
3334	Primary Aluminum	8	-	-	132 (125)	1 (1)	17 (16)	150 (142)	151 (143)
3221	Glass Containers	9	-	-	127 (120)	9.5 (9)	-- --	136 (129)	137 (130)
3352	Aluminum Rolling and Drawing	10	-	-	64 (61)	2 (2)	3 (3)	70 (66)	76 (72)
2821	Plastics and Resins	-	7	10	57 (54)	30 (29)	39 (37)	127 (120)	141 (134)
2611	Pulpmills	-	6	-	40 (38)	40 (38)	5 (4)	84 (80)	85 (81)
2815	Cyclic Intermediate and Crudes	-	8	-	14 (13)	29 (28)	76 (72)	119 (113)	133 (126)
2824	Organic Fibers (non-cellulosic)	-	9	8	44 (42)	27 (26)	63 (60)	135 (128)	139 (132)
2812	Alkalies and Chlorine	-	-	6	49 (47)	5 (4)	84 (80)	138 (131)	147 (139)
2823	Man-made Fibers (cellulosic)	-	-	7	13 (12)	2 (2)	67 (64)	82 (78)	84 (80)
3313	Electro-Metallurgical Products	-	-	9	2 (2)	1 (1)	54 (51)	57 (54)	59 (56)

(a) Gas at 40.9 MJ/Nm^3 (1038 Btu/scf), oil at 39.8 MJ/litre (6.0×10^6 Btu/42 gallon barrel), coal at 30.5×10^3 MJ/metric ton (26.2×10^6 Btu/ton).

(b) Total of purchased fossil (gas, oil, coal).

(c) Total of all purchased fuels and purchased electric power.

TABLE 4. RANKING OF INDUSTRY GROUPS BY SELECTION CRITERIA
(High numbers indicate higher ranking)

Selection Criteria	Petroleum Refining SIC 2911	Blast Furnaces and Steel Mills SIC 3312	Industrial Organic and Inorganic Chemicals SIC 2818,2819	Hydraulic Cement SIC 3241	Paper and Paperboard Mills SIC 2621,2631	Primary Aluminum SIC 3334	Glass Containers SIC 3221
(1) Industry Consumption of Natural Gas and Oil	7	5	6	3	4	1	2
(2) Amenability of Processes to Low- and Intermediate-Energy Gas	7	6	1	4	5	3	2
(3) Industry Location Relative to Coal Supplies	4	7	3	5	2	1	6
(4) Industry Dependence on Natural Gas and Oil	4	6	7	1	2	3	5
(5) Potential for Long-Term Utilization of Low- and Intermediate-Energy Gas	6	7	1	2	3	4	5
Total	28	31	18	15	16	12	20

Petroleum Refining (SIC 2911)

The petroleum refining industry ranks first in use of natural gas and fourth in use of oil for fuel among American industries ranking it very high under Criterion 1. The use of natural gas in this industry is about twice that of the second highest user, blast furnaces and steel.

The major uses of fuel in a refinery are for process feed stock heating and in boilers for steam generation. The relative amounts of steam and fuel for energy for various refinery processes are given later in this report (Table 16).

Process heaters are similar to water tube boilers in design. Fuel is fired into the heater which contains banks of tubes through which the oil is pumped. The major fuels used for this purpose are natural gas, refinery gas, and possible some residual oil. These process heaters and boilers are considered relatively amenable to being converted to a low- or intermediate-energy gas from coal(4). Therefore, refineries also rank high under Criterion 2.

The location of an industry relative to available supplies of coal is also important because on-site gasification would require shipment of coal to the plant. Table 5 gives installed refinery capacity in various states having significant coal resources.

The first segment of Table 5 lists refinery capacity in the top 10 coal producing states. This installed capacity accounts for 23.5 percent of total U.S. refinery capacity. The second segment of Table 5 includes refineries that have significant coal reserves although not necessarily high production rates. When this segment is added to that for high coal-producing states, the total installed capacity accounts for 58.3 percent of the total U.S. capacity. This segment of the industry would still rank first in natural gas use and sixth in oil use of those industries shown in Table 1. Therefore, the refinery industry ranks high under Criterion 3.

Refinery gas consists of off gases from various processes in the refinery. This gas can be a high grade fuel gas consisting of hydrogen, methane, ethane, propane, butane, and possible other hydrocarbons. In 1973, refinery gas accounted for about one-third of energy consumed in U.S. refineries and was exceeded only by natural gas in fuels consumed for energy. By using refinery gas, residual oil, and its crude feed for energy, if necessary, a refinery could always be energy self-sufficient. However, many of these items,

TABLE 5. TARGET INDUSTRY LOCATION IN RELATION TO COAL AVAILABILITY FOR REFINERIES

Rank	Coal Producing States	Refining Capacity, 10^6 litre/day	Refining Capacity, barrel/day
1	Kentucky	26.1	164,000
2	West Virginia	3.13	19,750
3	Pennsylvania	120.3	757,020
4	Illinois	185.7	1,168,150
5	Ohio	93.7	589,770
6	Virginia	8.42	53,000
7	Indiana	89.5	563,275
8	Alabama	5.46	34,375
9	Tennessee	6.98	43,900
10	New Mexico	16.4	103,060
		555.8	3,496,300

Rank	Coal Bearing States	Refining Capacity, 10^6 litre/day	Refining Capacity, barrel/day
4	Alaska	10.5	66,050
7	Colorado	9.5	60,000
15	Kansas	71.1	447,180
14	Missouri	17.1	107,000
2	Montana	25.0	157,206
1	North Dakota	9.32	58,658
17	Texas	624.7	3,929,430
13	Utah	22.7	143,000
5	Wyomin g	29.7	186,870
		819.5	5,155,394
Total		1,375.3	8,651,694
Total (all industry)		2,360.0	14,845,407

Percent of Total in

Coal Producing States	23.5
Coal Producing and Coal-Bearing States	58.3

particularly the various constituents in refinery gas, can be recovered as premium fuels with wider adaptability than low- or intermediate-energy gas from coal. Therefore, although refineries have alternatives to natural gas as a fuel, it may be highly attractive if these premium fuels resulting from the refining operations are released for other uses and low- or intermediate-energy gas from coal is fired in their place. This would result in refineries being relatively attractive under Criterion 4.

The petroleum industry is also energy intensive. The gross energy consumed per 1967 dollar of value added for 1971 was 300 MJ (284×10^3 Btu) per 1967 dollar(6). This value is higher (indicating more energy intensive) than any other industry considered in this study. Also, refineries represent considerable capital investment and installing a high investment long-term energy supply system, such as a coal gasification plant, would be within reason. Refineries would, therefore, tend to rank high under Criterion 5.

Blast Furnaces and Steel Mills (SIC 3312)

In conversations with representatives of the iron and steel industry, it was discovered that, for purposes of evaluating the applicability of gas from coal, basic steel mills (as represented by SIC 3312) should be generically separated into integrated and nonintegrated types. Integrated mills are generally large plants that produce semifinished steel products directly from iron ore. Secondary or nonintegrated mills tend to be smaller plants and produce a somewhat more finished grade of product but start with iron and steel scrap or prereduced iron ore pellets.

The principle differences in these two generic types of plants relative to the potential for utilization of low- or intermediate-energy gas is in the ability of the integrated plant to reduce iron ore to iron in a blast furnace by combining the raw ore with coke and lime under intense heat. The blast furnace, along with the coke oven used to pyrolize coal into coke, produces by-product fuels in the form of combustible tars and gases which reduce the need for higher grade fuels such as oil and natural gas. Secondary mills, on the other hand, have no sources of such by-product fuels and must rely on purchased fuels such as oil and natural gas as their energy sources. A recent AISI survey⁽⁷⁾, for example, revealed that in 1973 a sample of integrated steel plants was only about 18 percent dependent on natural gas as a source of energy, whereas a sample of secondary mills was 66 percent dependent.

Discussions with representatives of the integrated steel industy revealed that integrated plants would react to shortages of natural gas and oil by increasing their utilization of by-product blast furnace gas, coke-oven gas, and tars. Blast furnace gas with a heating value of about 3.5 MJ/Nm^3 (90 Btu/scf) can be used for underfiring coke ovens, boilers, and preheating air injected into the blast furnace. Coke-oven gas with a heating value of about 19.7 MJ/Nm^3 (500 Btu/scf) can be used almost anywhere in the plant where natural gas is used. Tars, which are normally sold, could be used as an additional source of fuel, if necessary.

In the secondary steel industry, there are few alternatives to natural gas as a fuel. Many processes in the industry require a clean gaseous or liquid fuel. In recent years gas shortages have forced plants to use more oil and propane, fuels which are expensive and occasionally hard to obtain due to short supply. Low- or intermediate-energy gas could, therefore, be an attractive fuel for this segment.

Statistics do not distinguish specifically between the integrated and secondary segments of the industry. Therefore, in order to determine how the secondary steel industry ranks under Criterion 1, the potential displacement of natural gas must be estimated from available data.

A 1973 survey of energy use in 16 sample integrated companies and 35 sample secondary companies⁽⁷⁾ produced the data given in Table 6. Assuming that the 16 integrated companies sampled were the largest of the 18 integrated companies, the figures shown in Table 6 represent 186 of 195 blast furnaces⁽⁸⁾ or roughly 95 percent of national capacity. The estimated total integrated steel production in 1973 would then be

80,321,430/0.95 = 84,548,873 metric tons
(93,321,000 tons).

TABLE 6. ENERGY USE PATTERNS IN SAMPLE INTEGRATED AND SECONDARY STEEL PLANTS (1973)[7]

Total 1973 Industry Shipments	100,978,230 metric tons (111,455,000 tons)	
For the 16 sample integrated companies		
Net shipments	80,321,430 metric tons (88,655,000 tons)	
Natural gas consumed	566×10^3 MJ	(537×10^{12} Btu) 6.0×10^6 Btu/ton
Oil consumed	241×10^3 MJ	(229×10^{12} Btu) 2.6×10^6 Btu/ton
Propane consumed	0.84×10^3 MJ	(0.8×10^{12} Btu) 2.6×10^6 Btu/ton
For the 35 sample secondary companies		
Net shipments	6,220,596 metric tons (6,866,000 tons)	
Natural gas consumed	73×10^3 MJ	(69.5×10^{12} Btu) 10.1×10^6 Btu/ton
Oil consumed	10×10^3 MJ	(9.6×10^{12} Btu) 1.4×10^6 Btu/ton
Propane consumed	0.3×10^3 MJ	(0.3×10^{12} Btu) 1.4×10^6 Btu/ton

Assuming also that the 35 secondary companies sampled were chosen more or less randomly from the 126 companies (listed in Reference 9), then the figures in Table 6 would represent approximately 35/126 or 28 percent of the national capacity. The estimated total secondary steel production in 1973 would the be

6,220,596/0.28 = 22,216,414 metric tons
(24,521,428 tons).

The total production estimated by this method is 106,762,240 metric tons (117,842,428 tons)(9)

By this estimation, integrated steel plants account for roughly 79 percent of steel production and secondary steel plants approximately 21 percent. Applying the energy use per ton data of Table 6, the resulting energy use estimated in Table 7 are obtained.

It is estimated that the secondary steel industry consumes approximately 31 percent of the total natural gas consumed in the steel industry and 26 percent of the combined natural gas and oil. In 1974, the latest year for which figures are available, the total natural gas used by the steel industry was 704.6 x 10^9 MJ (667.9 x 10^{12} Btu). Therefore, considering conversion of the secondary steel industry to fuel gas from coal, the estimated possible displacement of natural gas is 218.4 x 10^9 MJ (207 x 10^{12} Btu), and the displacement of oil and natural gas combined is 288 x 10^9 MJ (273 x 10^{12} Btu)*.

In 1971, the year for which industry fuel-use data are presented in Table 3, the secondary steel industry would have consumed about 213.1 x 10^9 MJ (202 x 10^{12} Btu) of natural gas and about 21.7 x 10^9 MJ (20.6 x 10^{12} Btu) of oil. This would rank the secondary steel industry eighth in natural gas use and also eighth in combined use of natural gas and oil. This would still rank the secondary steel industry high under Criterion 1.

Fuel-using processes in the secondary steel industry are similar to those in the primary or integrated industry and consist of various types of furnaces for heating steel for a variety of forming and heat-treating

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*1974 was not a high-production year for the steel industry; thus, these values might be considered minimum values.

TABLE 7. ESTIMATED ENERGY USE PATTERNS IN THE INTEGRATED AND SECONDARY STEEL INDUSTRY

Estimated integrated steel company data		
Net shipments	84,548,826 metric tons (93,321,000 tons)	
Natural gas consumed	591×10^9 MJ	(559.93×10^{12} Btu)
Oil and propane consumed	256×10^9 MJ	(242.63×10^{12} Btu)
Estimated secondary steel company data		
Net shipments	22,216,414 metric tons (24,521,428 tons)	
Natural gas consumed	261×10^9 MJ	(247.67×10^{12} Btu)
Oil and propane consumed	36×10^9 MJ	(34.3×10^{12} Btu)
Total Estimated Natural Gas Use	851×10^9 MJ	(807×10^{12} Btu)
Total Estimated Oil and Propane Use	292×10^9 MJ	(277×10^{12} Btu)
Percent of total industry natural gas used by secondary mills	31 percent	
Percent of total industry natural gas, oil, and propane used by secondary mills	26 percent	

operations. Many of these furnaces, like those in the integrated steel industry, are designed to allow conversion from natural gas to coke-oven gas which has a heating value of about 19.7 MJ/Nm^3 (500 Btu/scf). Many of the furnaces are considered relatively easy to convert to at least an intermediate-energy gas with 11.8 MJ (300 Btu/scf). Thus, the secondary steel industry ranks relatively high under Criterion 2.

A further refinement of the potential for use of low- and intermediate-Btu gas in the secondary steel industry is obtained by estimating that segment of the industry in relative close proximity to sources of coal. Table 8 lists installed electric-arc furnace capacity (electric arc furnaces are used nearly exclusively in the secondary steel industry) in the top 10 coal-producing states(10) and in 9 additional states that have significant coal reserves. Thus, 57.6 percent of the industry is located in the top 10 coal-producing states, and 70.4 percent of the industry is located in states having significant production or significant reserves. Considering that the secondary steel industry located in coal-producing or coal-bearing states has a high potential for conversion, a more refined estimate of the amount of natural gas and fuel oil that could be displaced by low- or intermediate-energy gas in the secondary steel industry is approximately 89.6×10^9 to 105.5×10^9 MJ (85×10^{12} to 100×10^{12} Btu) of natural gas and from 100.2×10^9 to 122×10^9 MJ (95×10^{12} to 116×10^{12} Btu) of natural gas and fuel oil per year. Thus, in relation to Criterion 3, the secondary iron and steel industry would rank high.

The dependence of the secondary steel industry on a source of clean gaseous or liquid fuel was mentioned earlier. The major portion of fuel use in a secondary mill is for furnaces which generally could not be converted to direct firing of fuels such as coal, nor could these processes be easily replaced with those that could fire coal directly. Thus, the secondary steel industry would rank high under Criterion 4.

The basic metals industries are considered highly energy intensive. The gross energy consumption per 1967 dollar of value added for SIC 331

TABLE 8. TARGET INDUSTRY LOCATION IN RELATION TO COAL AVAILABILITY FOR SECONDARY STEEL MILLS

Rank	Coal Producing States	Electric Arc Capacity, 10^3 metric ton/year	Electric Arc Capacity, 10^3 ton/year
1	Kentucky	625	690
2	West Virginia	136	150
3	Pennsylvania	7,234	7,985
4	Illinois	4,222	4,660
5	Ohio	3,420	3,775
6	Virginia	245	270
7	Indiana	1,110	1,225
8	Alabama	489	540
9	Tennessee	226	250
10	New Mexico	--	--
		17,708	19,545

Rank	Coal Bearing States	Electric Arc Capacity, 10^3 metric ton/year	Electric Arc Capacity, 10^3 ton/year
4	Alaska	--	--
7	Colorado	453	500
15	Kansas	45	50
14	Missouri	1,087	1,200
2	Montana	--	--
1	North Dakota	--	--
17	Texas	2,346	2,590
13	Utah	--	--
5	Wyoming	--	--
		3,932	4.340
Total		21,640	23,885
Total (all industry)		30,743	33,933

Percent of Total in

Coal Producing States	57.6
Coal Producing and Coal Bearing States	70.4

(basic metals) was 44.7 x 10^3 kcal (177.5 x 10^3 Btu) per 1967 dollar of value added(6) for 1971 which is one of the highest of the various industries considered in this study. This would result in secondary steel also ranking high under Criterion 5.

Other Industries Considered

Seven other industrial groups that consume significant amounts of natural gas and oil for fuel were also considered along with petroleum refining and secondary steel mills. These seven other groups were felt to be less attractive than the two selected; however, they would not in all cases be considered unattractive for study.

Industrial Organic and Inorganic Chemicals [not elsewhere classified] (SIC 2818, 2819). The industrial chemical industry (SIC 281) involves the production of a wide variety of chemical products. The two groupings considered here, organic and inorganic chemicals not elsewhere classified, are by far the major fuel-using groups within the industrial chemicals industry; however, they also involve a wide variety of products.

In production of chemical products, the major use of natural gas and oil as fuel is in boilers for steam generation. In one study(11), boiler fuel use was estimated at 50 to 60 percent of total process energy needs, and in discussions with a major chemical company, it was learned that boiler fuel use in many plants could range from 75 to 90 percent of all process energy use. In discussions with this same company, it was learned that it would be more economical to replace existing gas- and oil-fired boilers with coal-fired boilers equipped with appropriate pollution control devices (particulate and SO_2) than to build a coal gasification plant for firing coal.

Nonboiler fuel use in a chemical plant consists of items such as natural gas reforming, catalyst regeneration, and feedstock heating or heating to maintain prescribed reaction temperatures. Due to the combustion requirements of these processes (extremely close temperature control with a large number of burners per unit), thery are considered unattractive to retrofitting to low- or intermediate-energy gas from coal.

For these reasons the industrial chemical industry as a whole was considered unattractive for use as a target industry in this study. This, however, should not imply that in particular cases industrial chemical plants may not be attractive applications for low- and intermediate-energy gas from coal.

Hydraulic Cement (SIC 324). The manufacture of hydraulic cement was also considered as a potential target industry. Cement plant fuel use is confined almost entirely to one process; firing the long inclined rotating kilns used in producing the clinker necessary for making cement. These kilns are commonly fired from the product discharge end with a small number of large burners.

Many cement plants are fired with coal, and in fact, the hydraulic cement industry ranks first in coal consumption for U.S. industry (Table 3). Kilns fired with oil or natural gas would be considered relatively convertible to low- or intermediate-energy gas. However, because the cement industry is relatively evenly scattered throughout the country to maintain close proximity to markets and minimize shipping costs, many of those plants designed for natural gas and oil would be located in areas where coal is not readily available. Also, if coal were available, it may be more economical to convert a gas- or oil-fired kiln to direct coal firing. For these reasons the cement industry, though potentially attractive for application of low- and intermediate-energy gas, was not selected as a target industry for this study.

Paper and Paperboard Mills (SIC 2621, 2631). This industry combination ranks fourth in combined consumption of natural gas and oil of those industries listed in Table 3. Industries in this group were not selected as target industries because the vast majority of fuel used in a paper mill is for boiler applications. Installation of a coal gasification plant to supply gas primarily for boiler firing is not considered economically competitive with direct coal-fired boilers. Also, a large fraction of the paper industry is located in the northwest and northeast sections of the country--close to supplies of timber but where coal availability is low.

Primary Aluminum Industry (SIC 3334). Primary aluminum ranks eighth in use of natural gas for industries listed in Table 3. This fuel is used principally for calcining alumina and refining and melting operations in the plant as the basic production process uses the Hall Heroult electrolytic method. Due to the high electric energy required for producing aluminum, most aluminum plants are located in areas where cheap hydroelectric power is readily available, such as the northeast and northwest area of the United States. These areas generally have poor coal availability and, thus, primary aluminum was not considered as attractive as the two target industries selected.

Glass Containers (SIC 3221). The glass container industry is considered a relatively attractive candidate for conversion to intermediate-energy gas. The glass melting operation is by far the major energy-consuming process, consuming about 85 percent of energy used in glass making. The melter consists of a large refractory-lined tank where usually natural gas is fired over the molten glass. Temperatures in the melter are high (often over 3000 F) and require close control. In natural-gas-fired melters, regenerators are used to preheat combustion air to obtain the necessary melter tempereatures and improve efficiency. Converting these melters to low-energy gas from air-blown gasifiers would be difficult. However, the higher flame temperatures and other properties of intermediate-energy gas make it an attractive candidate for substitution.

The glass industry is highly dependent on natural gas; however, many container glass plants (over 50 percent) use electric heating to boost the capacity of gas-fired melters. Some all-electric melters have been built and offer certain advanatages in lack of pollution-control requirements and higher thermal efficiency of the melter itself. In the future the glass industry could shift more and more to electric boosting and all-electric melting to relieve the dependence on natural gas(12)

As a result of these considerations, the glass industry was considered somewhat less attractive than the two target industries selected for this project.

II. GASIFIER AND GAS CLEANUP SELECTION

For each of the hypothetical model plants studied, a variety of gasifier-gas cleanup system combinations were considered. These various combinations were screened and eventually one combination was selected for each of the two model plants. During the screening, an attempt was made to appropriately match the gasification system characteristic with model plant requirements. Thus, the gasifier/plant combinations reported here appear to be reasonable selections for the model plants of interest. Actually, for most applications several different gasifier-gas cleanup systems might be selected for a given application by another investigator or after more in-depth analysis.

Gasifier Selection

In selecting gasifiers for model plants, attention was given only to those processes that had been commercially proven in the past and are currently commercially available. Table 9 lists the 10 processes considered along with some of their salient characteristics. These processes consist of three basic types: entrained slagging, fixed bed, and fluidized bed. Both air-blown and oxygen-blown configurations were considered.

Entrained slagging gasifiers are represented by the Koppers-Totzek and Babcock & Wilcox. Entrained slagging gasifiers have the advantage of being able to fire nearly any coal regardless of characteristics. They also are capable of being produced in larger unit sizes with higher gas production rates per unit. The gas produced in these units, due to its high temperature, usually contains no tars, phenols, or oils, and is also generally very low in ammonia.

Entrained gasifiers require pulverizing of the coal, usually to a size specified as 70 percent through a 200-mesh sieve. They also are primarily restricted to oxygen-blown operation, producing a gas of about 300 Btu/scf. The lower temperatures that accompany air-blown operation make it difficult to maintain slagging conditions, and when air-blown operation is possible, it produces a generally low-grade gas of around 4.7 MJ/Nm^3 (120 Btu/scf). In the oxygen-blown configuration, entrained slagging units generally require more oxygen

TABLE 9. COMMERCIAL GASIFIERS CONSIDERED FOR MODEL INDUSTRY PLANTS

Gasifier		Unit Output,		Potential Byproducts	Limitations
		10^6 MJ/day	(10^9 Btu/day)		
Lurgi	fixed/agitator	8.4-12.7	(8-12)	tar, oil, phenols, NH_3	Needs sized low-caking coal
Koppers-Totzek	entrained slagging	8.4-17.4	(8-16.5)	steam	Oxygen-blown only, pulverized coal
Wellman Galusha	fixed/agitator	1.6-2.6	(1.5-2.5)	tar, oil, phenols, NH_3	Needs sized low-caking coal
Applied Technology	fixed/2-stage	0.2-2.4	(0.2-2.3)	tar, oil, phenols, NH_3	Needs sized coal, free swelling index <3
Winkler	fluidized	1.4-14.8	(1.3-14)	steam	Needs crushed low-caking coal
Riley Morgan	fixed/agitator	1.6-3.2	(1.5-3.0)	tar, oil, phenols, NH_3	Needs sized low-caking coal
Woodall Duckham	fixed/2-stage	1.1	(1.0)	tar, oil, phenols, NH_3	Needs sized coal, free swelling index <3
Babcock & Wilcox	entrained slagging	16	(15)	steam	Primarily O_2-blown, pulverized coal
M. W. Kellogg	fixed/agitator	2.6-3.7	(2.5-3.5)	tar, oil, phenols, NH_3	Needs sized low-caking coal
Wilputte	fixed/agitator	0.7	0.6	tar, oil, phenols, NH_3	Needs sized low-caking coal

than fixed-bed or fluid-bed units. These units operate at high gas temperatures usually over 1650 C (3000 F) which could result in significant loss of sensible heat in the gas if extensive heat recovery equipment is not employed. These high gas temperaures do allow the entrained slagging unit to produce a significant amount of steam which may become a valuable by-product in some industry situations.

Fixed-bed gasifiers have certain limitations of coal-feed characteristics for proper operation. Generally, coals with low-caking characteristics or low free-swelling index (generally Western lignitic coal or some low-swelling, free-burning Eastern coals) work best in these gasifiers. Agitator-type fixed-bed units such as Lurgi, Wellman Galusha, Riley Morgan, Kellogg, and Wilputte can generally utilize coals with mildly caking characteristics--up to a free-swelling index of about 7. Two-stage fixed-bed units such as Applied Technology Incandescent and Woodall Duckham are limited to coals with a free-swelling index of less than 3. Fixed-bed units also require the coal to be carefully sized for proper operation. Coal feed is usually double screened to reduce the percentage of undersized and oversized particles. With bituminous coal, elimination of as great a percentage of the fines (particles less than 0.6 mm (1/4 inch)) as practical is important.

Fixed-bed units, due to their low temperature of operation, produce a significant amount of tars, phenols, and oils in the product gas. These must be removed prior to use in most industrial situations. These constituents may, in many cases, represent pollutants or potential wastes, but in some industrial situations may be used as by-products in the industrial processes. Fixed-bed units are also limited to relatively small unit capacities. This is due to an inherent limitation on the through-flow velocity and also a limitation on manufacturing large rotating parts (such as grates or agitator arms for the inside of a vessel). Fixed-bed units are capable of high thermal efficiencies, however, approaching 90 percent on a hot gas basis. They also have high turndown ratios and are capable of operating at very low loads in an efficient manner. Many fixed-bed units are rated at turndown ratios of over 90 percent.

Only one fluidized-bed gasifier, the Winkler, has been commercially proven at this time. In fact, the Winkler gasifier was the first demonstration of the fluidized-bed principle. Fluidized-bed gasifiers are limited to mildly caking coal, similar to fixed-bed gasifiers. Coal for the Winkler gasifier must first be crushed to a size less than 0.95 cm (3/8 inch), and undersize particles or fines can be tolerated. However, a significant amount of the ash (up to 70 percent) is carried away with the gas stream and must be removed prior to the gas being used. These units operate at gas temperatures of 815 to 982 C (1500 to 1800 F), about half way between those of fixed-bed units and entrained slagging units. These units are available in relatively large unit capacities, have relatively high turndown ratios, can operate at up to 50 percent over design capacity, and relatively simple in operation.

Industry Consideration in Gasifier Selection

The characteristics of a particular gasification system are important when considering its applicability to a particular industry. In some industries the production of tars, phenols, oils, and ammonia commonly produced in fixed-bed processes would be considered valuable byproducts. In other industries, however, these constituents would be troublesome wastes and would have to be disposed of in some acceptable manner. On the other hand, some industries may have a need for the large amounts of steam generated in cooling down the hot gas entrained slagging processes. Some industries may have the need for the high turndown capability and flexibility of fixed-bed units, whereas other industries may prefer the high on-stream factors and continuous operating capability of the entrained or fluid-bed units. Also, some industries will have the need for a higher grade gas--such as an intermediate-energy gas produced from an oxygen-blown gasifier. In these cases either an entrained, fixed-bed, or fluid-bed unit would be appropriate. In other cases, however, this higher grade gas may not be necessary, and an air-blown gasifier (without the necessity of an oxygen plant) would be satisfactory. These cases would tend to be selective for fixed-bed units which produce a higher grade gas when air-blown than either an entrained or fluid-bed unit.

The size or unit capacity of the gasifier is important when considering it for industry. Entrained and fluidized-bed units are offered primarily in large unit capacities. Fixed-bed units, on the other hand, are restricted to relatively small unit capacities. In a particular design, multiple numbers of gasifiers are attractive in that they offer greater load flexibility and the ability to carry some fraction of the load should one unit fail. Too many units, however, minimize the advantages of economy of scale and generate increased complexity. Hence, industrial plants with large energy demands may find it attractive to select a few entrained or fluidized-bed units; whereas, industrial plants with smaller energy demands will find it attractive to select several atmospheric fixed-bed units. These considerations were used in making a cursory evaluation of which gasifier, cleanup system combinations were most applicable to the two model plant studied.

Gasification Systems Selection for the Steel Plant Model

The steel plant model in this study has an energy demand of from 16 x 10^6 to 24 x 10^6 MJ/day (15 x 10^9 to 23 x 10^9 Btu/day). Therefore, only those gasifiers with larger unit capacities--that is, Koppers-Totzek, Babcock & Wilcox, Winkler, and Lurgi--were considered. The other units listed in Table 9 would involve a large number of individual units to satisfy this energy demand and were considered unattractive for this reason. The coal selected for use in the model steel plant was a lignite type with a low free-swelling index which would present no particular problem to any of the listed gasifiers. Therefore, feed stock characteristics were not considered important as a selection parameter in choosing a suitable gasification process.

The steel plant would have no use for byproducts such as tars, phenols, oils, or ammonia, and, therefore, attention was directed primarily at entrained and fluidized-bed processes which do not produce significant amounts of these products. On the other hand, the steel plant could possibly have use for a low-grade steam, either for space heating or for process use, which would make the selection of entrained or fluidized-bed processes more attractive. Of the three nonfixed bed processes (Koppers-Totzek, Babcock &

Wilcox, and Winkler), the Koppers-Totzek was chosen as the representative gasification process for detailed consideration. This selection was based primarily on the fact that more information was available on the performance of the Koppers-Totzek process process than either of the other two gasifiers. Both the Babcock & Wilcox and Winkler processes are also attractive, and designs based on these processes should involve similar considerations to that based on the Koppers-Totzek.

Gasification Systems Selection for the Refinery Model

The model refinery has an energy demand approximately 3.5 x 10^6 to 4.6 x 10^6 MJ/day (3.3 to 10^9 to 4.4 x 10^9 Btu/day). Due to this relatively small energy demand, fixed-bed units with lower unit capacities were considered attractive for selection. Also, because the gas made from coal would be blended with refinery gas having a heating value of about 39.4 MJ/Nm^3 (1000 Btu/scf), air-blown operation producing low-energy gas was considered to be satisfactory. Also, the refinery potentially could utilize the tars, phenols, and the other chemical products in fixed-bed processes.

The coal selected for the refinery was a lignitic coal having a free-swelling index of about 4 to 4-1/2. This restricted the selection of a gasification process to those fixed-bed units having agitator-type beds. The two-stage fixed-bed processes such as Woodall-Duckham and Applied Technology Incandescent are restricted primarily to coals with free-swelling index of less than 3.

The Wellman Galusha was selected over the other small agitator-type fixed-bed units of Riley Morgan, Kellogg, and Willputte, based primarily on the fact that in the recent past it has achieved a greater degree of commercial application. Two gasification plants currently are still operating using Wellman Galusha gas producers, and the information on operational characteristics provided from these plants was considered to be potentially useful in the study. The other agitator-type fixed-bed units would also be attractive for consideration and results obtained from analysis with the Wellman Galusha would be expected to be similar to that for the other gasifiers. If a suitable supply of coal with free-swelling index less that 3 could be obtained, consideration of two-stage fixed-bed units would provide an interesting comparison to that of the agitator-fixed bed.

Selection of Gas Cleanup Systems

A wide variety of cleaning systems are available for removing sulfur compounds from fuel gas. All currently commercially available system operate at low temperatures below 120 C (250 F) and would require precleaning the gas of particulates, tars, and other constituents that may interfere with the gas desulfurization step along with cooling to the prescribed operating temperature.

Most commercially available desulfurization systems use wet scrubbing for removing sulfur compounds from the fuel gas. These systems can generically be separated into physical, chemical, and physical-chemical absorption/desorption types. A special class of chemical absorption sytems involves direct oxidation of H_2S to sulfur rather than desorption.

Physical absorption processes normally operate at higher pressures and are capable of reducing H_2S, COS, and CS_2 to extremely low levels. These processes can also be made selective for H_2S over CO_2, producing a concentrated H_2S steam suitable for sulfur recovery in a Claus unit. Physical sorbent processes, however, have a high solubility of hydrocarbons in the sorbent and sorbent costs are high. Examples of physical sorbent processes are Lurgi Purisol and Rectisol, Allied Chemical Selexol, and Fluor solvent.

Chemical absorption processes include scrubbing with ammonia or alkali solutions. Mono amine systems have high sorbtivities for H_2S but are sensitive to deactivation from reaction with COS. Di-amines and tertiary-amines are less sensitive to deactivation but have lower solubilities of H_2S and remove little or no COS or CS_2. Amine systems also remove increasing amounts of CO_2 as the sulfur level in the fuel gas is reduced. Alkaline salt sorbents remove both sulfur compounds and CO_2. Examples of chemical sorbent processes are monoethanolamine (MEA), diethanolamine (DEA), and monoethanol-diethanolamine (MDEA).

Physical-chemical desulfurization processes use a combination of physical and chemical sorbent. These processes attempt to take maximum advantage of the attractive characteristics of both physical and chemical types of processes while minimizing the unattractive features of each. Examples of these processes are Shell Sulfinol and Lurgi Amisol.

Some types of processes such as chemical sorbent systems also remove CO_2 in addition to H_2S. Removal of CO_2 makes it difficult to obtain a sufficiently high concentration of H_2S (at least 15 percent) in the gas resulting from sorbent regeneration to use a Claus plant for sulfur recovery. In these cases, it may be necessary to use a chemical absorption/direct oxidation process for sulfur recovery. The direct oxidation processes can also be used directly on the fuel gas for removal of sulfur compounds. These processes involve absorption of sulfur compounds to elemental sulfur. Examples of direct oxidation processes are Stretford and Giammarco-Vetrocoke.

In addition to wet desulfurization systems various dry removal systems have been used commercially over the years. These processes operate at low temperatures and involve adsorption of sulfur compounds on iron oxide (supported by various media), activated carbon, or molecular sieves. Although some of these processes were used at one time for H_2S removal from producer gas and coke-oven gas, they are not generally considered applicable to low- or intermediate-energy gas today due to requirements for large amounts of sorbent, high cost of sorbent, or necessity for maintaining prescribed humidity and gas temperature for proper operation.

The commercial fuel-gas desulfurization systems exhibit a wide variety of operating characteristics. Optimizing the selection of any one of these processes would necessarily involve many detailed engineering considerations. For this study, systems were selected that would be likely candidates for an actual plant design based on gasifier vendor's recommendations. For the steel plant model using a Koppers-Totzek gasifier, an MDEA system was chosen and for the refinery model, using a Wellman Galusha gasifier, a Stretford system was chosen. Both gasifier vendors recommend the use of these respective cleaning systems as a first consideration in a plant design similar to those evaluated in this study. Both vendors, however, may recommend different systems if warranted by a particular application.

Both the MDEA and Stretford systems involve chemical absorption. Both these systems can operate satisfactorily at the near ambient pressures of the Koppers and Wellman Galusha gasifiers. In the MDEA system, H_2S and up to 75 percent of the COS are absorbed in an amine solution which is subsequently regenerated with steam and pressure reduction yielding an H_2S - rich gas stream. The H_2S-rich gas stream is suitable for feed to a Claus unit which converts about 95 percent of the H_2S to elemental sulfur. In the Stretford process, H_2S is absorbed in Stretford solution (a dilute ammonium vanadate, sodium citrate, and soda ash), and this is oxidized in solution to elemental sulfur which is then filtered from solution.

III. CONVERSION OF A SECONDARY STEEL PLANT TO INTERMEDIATE-ENERGY GAS

The Secondary Steel Industry

The secondary steel industry produces semifinished or finished products from iron or steel scrap or prereduced iron ore pellets. Because the feedstock for a secondary steel plant is in a reduced or semi-refined state, blast furnaces and coke ovens (which are used in integrated plants to reduce iron ore to iron) are not required. The electric-arc furnace is almost exclusively the type of furnace employed in secondary steel mills for making steel from scrap or prereduced iron ore although some open-hearth furnaces remain in service. Integrated steel plants use primarily the basic oxygen furnace for steelmaking.

The electric-arc furnace is a short cylindrical shaped furnace having a rather shallow hearth. Three carbon, or graphite, electrodes project through the roof into the furnace. Charge materials consist of 100 percent scrap and the required alloying elements. Electric energy passing through the electrodes into the metallic charge creates the heat required to melt the charge. When the first scrap charge is almost completely melted a second and a third scrap charge may be added, depending on the size of the furnace and density of the scrap. The molten steel is poured into ingot molds, where the steel solidifies before further processing.

Solidified ingots are removed from the molds and placed in soaking pits where the temperature of the ingot is permitted to equalize, after which the temperature of the ingot is raised to the required temperature for rolling. In an electric-arc furnace steelmaking shop the soaking pits are fired with natural gas or fuel oil. The hot ingots are delivered to a blooming mill where they are rolled to slabs, blooms, or billers, depending on the size of the ingot and the end-products produced at any particular steel plant. An alternate method for producing these intermediate products is to use a continuous casting machine which makes the slabs, blooms, or billets, directly from the molten steel. In such a secondary steel making plant the soaking pits and blooming mill are not required thus reducing the overall plant energy requirements.

Slabs, blooms, and billets are permitted to cool to ambient temperatures, after which they are inspected, defects removed, and eventually they are reheated for additional rolling. The reheating furnaces are usually fired with natural gas or fuel oil. Some smaller reheating furnaces may be fired with propane. Reheated slabs are rolled into plate, sheet, or strip; reheated blooms into heavy structural beams, channels, and railroad rails; and reheated billets are rolled into angles, channels, reinforcing bar, round bar, square bar, rod, and other "merchant products".

Some electric furnace shops may manufacture finished products, in addition to the usual semifinished products. The finished products made may include wire fencing, reinforcing mesh, building joists, nails, and miscellaneous forged products, to name a few.

Industry Data

Secondary steel mills make a significant contribution to the total amount of steel produced in the United States. These plants produce plain-carbon, alloy, and stainless steels. Production statistics for secondary steel mills from 1965 through 1974 are given in Table 10.

The secondary steel industry consists of electric-arc furnace steelmaking plants which primarily use steel scrap for the complete melting charge. One electric-arc furnace steelmaking shop, that is associated with an integrated steelmaking plant, routinely uses molten pig iron for about 50 percent of the metallic charge. One or two other electric-arc furnace shops, in similar circumstances, will occasionally use molten pig iron for about 50 percent of the metallic charge.

TABLE 10. SECONDARY STEEL MILL PRODUCTION IN THE UNITED STATES(9)

Year	Carbon Steel, 10^6 metric tons	Carbon Steel, 10^6 tons	Alloy Steel, 10^6 metric tons	Alloy Steel, 10^6 tons	Stainless, 10^6 metric tons	Stainless, 10^6 tons	Total Arc-Furnace Steel, 10^6 metric tons	Total Arc-Furnace Steel, 10^6 tons	Percent of Total U. S. Steel Production
1965	7.6	(8.4)	3.5	(3.8)	1.35	(1.4)	12.5	(13.8)	10.5
1966	8.3	(9.1)	3.6	(4.0)	1.49	(1.6)	13.4	(14.8)	11.1
1967	9.1	(10.1)	3.2	(3.5)	1.34	(1.4)	13.6	(15.0)	11.9
1968	9.9	(10.9)	4.0	(4.3)	1.33	(1.4)	16.2	(16.8)	12.8
1969	12.1	(13.3)	4.8	(5.1)	1.40	(1.5)	18.1	(20.1)	14.3
1970	12.9	(14.1)	4.3	(4.7)	1.16	(1.2)	18.2	(20.1)	15.3
1971	13.5	(14.9)	4.3	(4.7)	1.15	(1.2)	18.3	(20.9)	17.4
1972	14.8	(16.4)	5.2	(5.7)	1.40	(1.5)	21.5	(23.7)	17.8
1973	17.3	(19.1)	6.1	(6.7)	1.71	(1.8)	25.1	(27.7)	18.4
1974	17.7	(19.5)	6.2	(6.9)	1.09	(2.1)	26.0	(28.6)	19.7

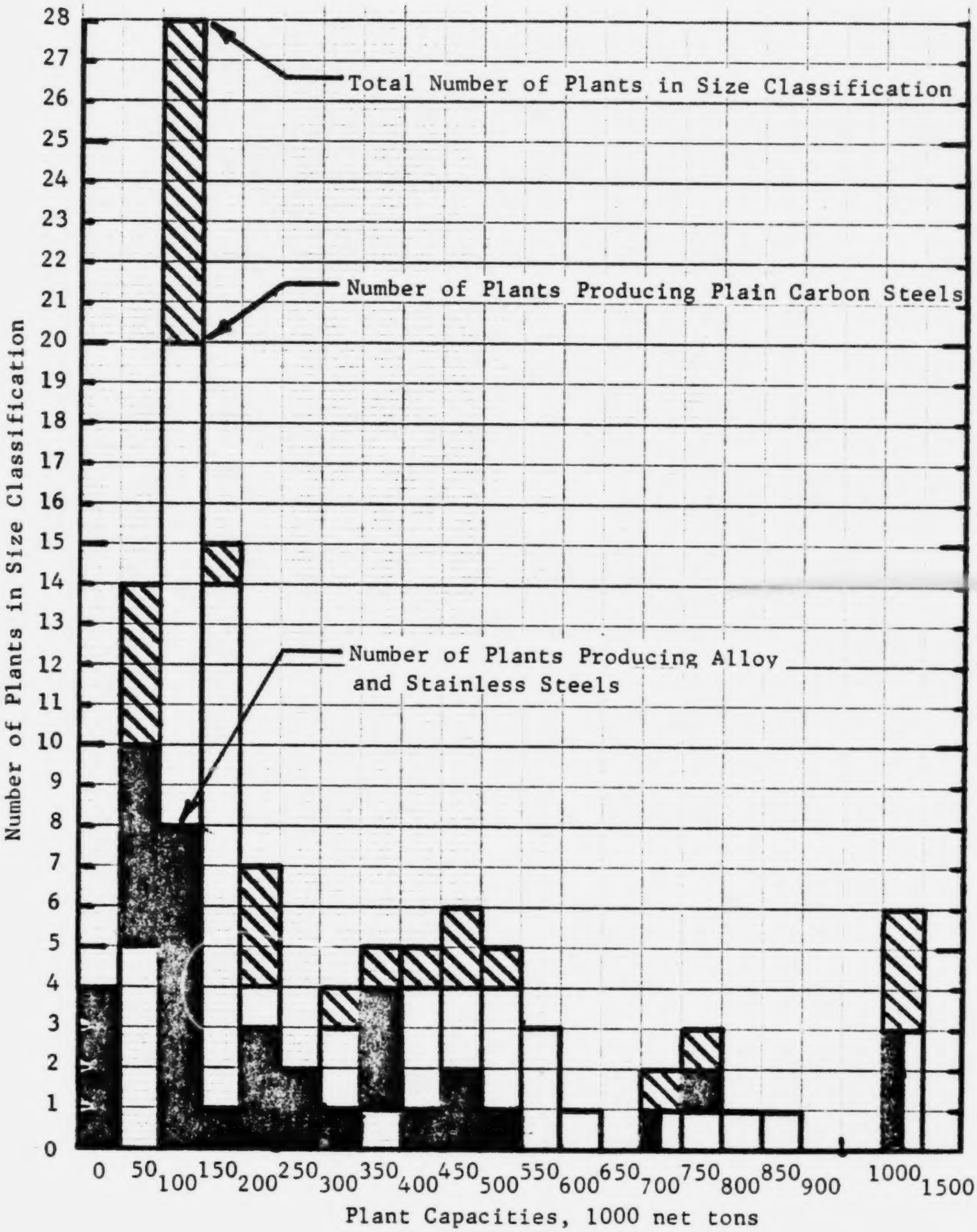
There are 126 electric-arc furnace steelmaking plants in the United States. Of these, 14 plants make steel for the manufacture of forgings and related products. The remaining 112 plants which are considered representative secondary steel plants for this study produce steel that is rolled into strip, sheet, bars, rods, angles, channels, and related products. Plain carbon steels are made in 70 plants, with the remaining 42 plants producing primarily alloy and stainless steels. A statistical distribution of the annual production capacities of the 112 arc-furnace steelmaking plants is shown in Figure 3.

Of the 112 electric-arc furance steelmaking plants that fall into the category for this study, 53 have continuous casting operations, leaving 59 plants that still operate with soaking pits and blooming mills. Figure 4 shows a statistical distribution of electric-arc furnace plants, according to annual steelmaking capacities, that use continuous casting machines to convert the molten steel into slabs, blooms, and billets. It should be noted that steelmaking plants with annual capacities of 220,800 metric tons (200,000 net tons) per year or less have a significant number of continuous casters.

The locations of the electric-arc furnace steelmaking plants are shown in Figure 5, superimposed on a map showing the coal fields of the United States.

Model Electric-Arc Furnace Steel Making Plant

It would be ideal to select a plant that could be characterized as being "typical" in its representation of the electric-arc furnace steelmaking plants in the United States. However, such is not the case, especially with



Note: 50 denotes melting plants producing from 50,000 to 99,000 net tons per year; 100 denotes melting plants producing from 100,000 to 149,000 net tons per year; etc.

FIGURE 3. STATISTICAL DISTRIBUTION OF THE CAPACITIES OF ELECTRIC-ARC FURNACE MELTING PLANTS IN THE UNITED STATES

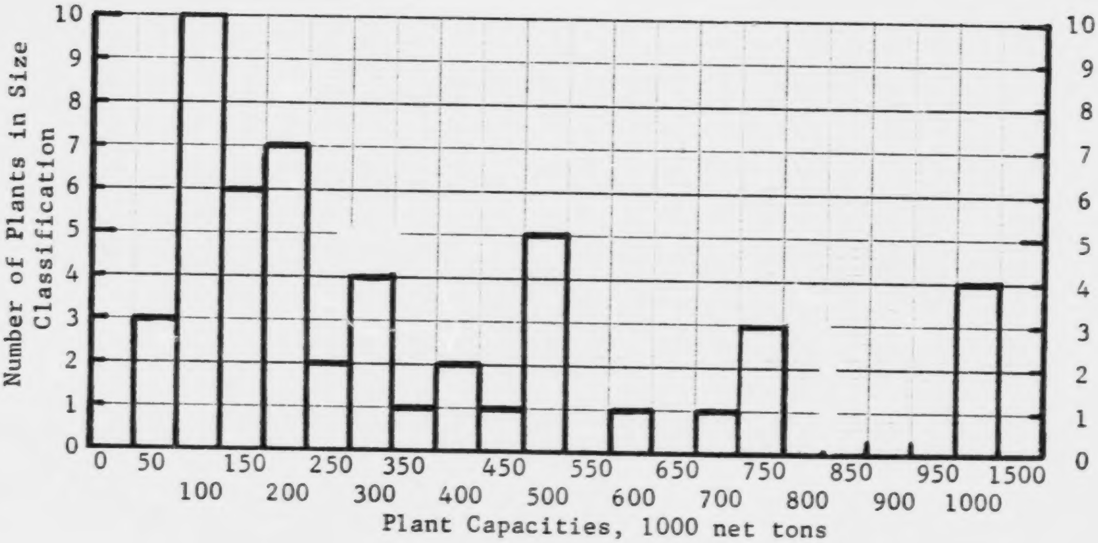
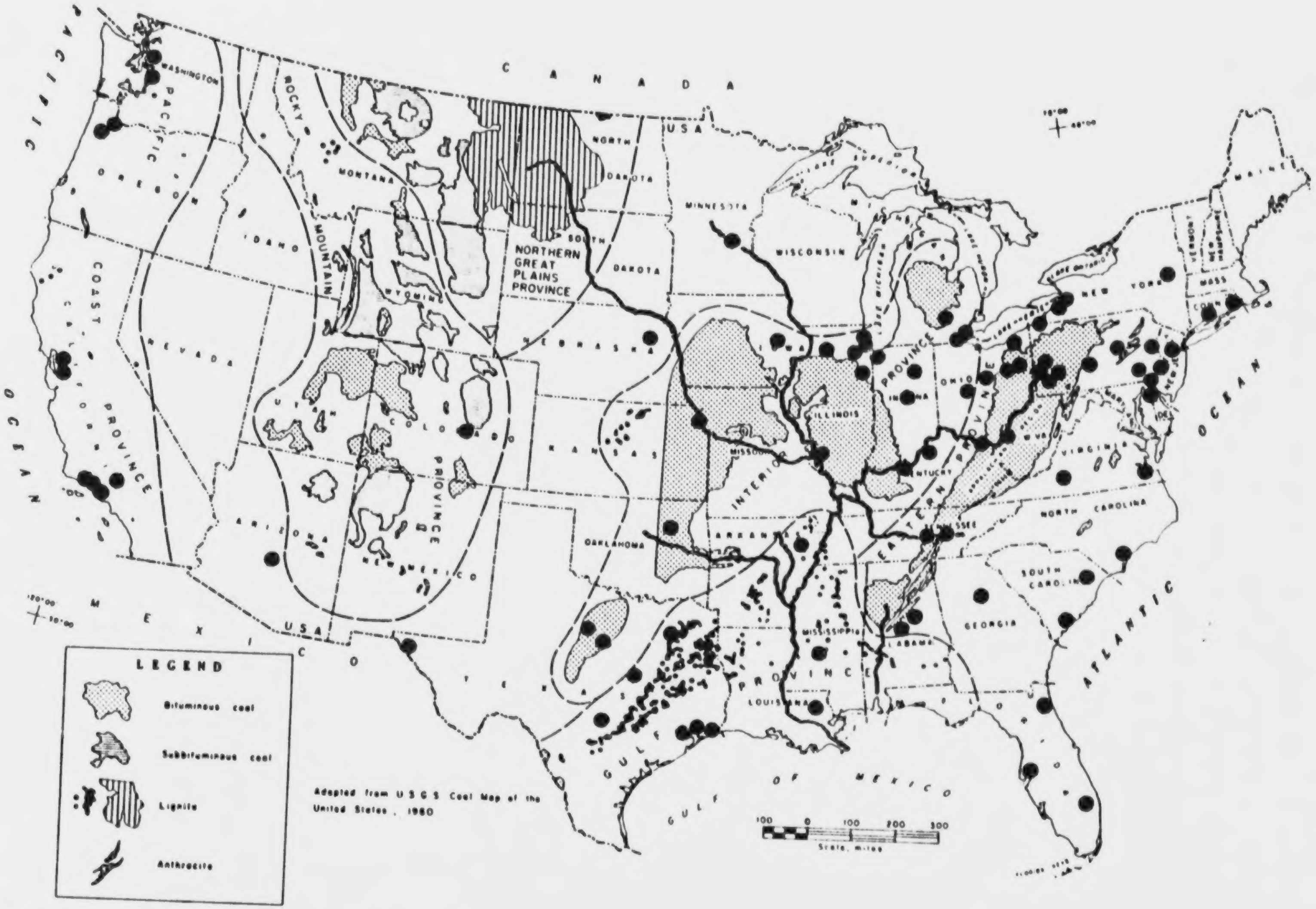


FIGURE 4. STATISTICAL DISTRIBUTION OF ELECTRIC-ARC FURNACE MELTING PLANTS IN THE UNITED STATES HAVING CONTINUOUS CASTING MACHINES


LEGEND
Bituminous coal
Subbituminous coal
Lignite
Anthracite
CANADA
MEXICO
PACIFIC OCEAN
ATLANTIC OCEAN
GULF OF MEXICO
PACIFIC COAST PROVINCE
ROCKY MOUNTAIN PROVINCE
NORTHERN GREAT PLAINS PROVINCE
INTERIOR PROVINCE
EASTERN PROVINCE
GULF PROVINCE
100 0 100 200 300
Scale, miles

Adapted from U S G S Coal Map of the United States, 1960

FIGURE 5. LOCATION OF ELECTRIC-ARC FURNACE STEELMAKING PLANTS AND THE LOCATION OF BITUMINOUS, SUBBITUMINOUS AND LIGNITE COAL FIELDS IN THE UNITED STATES

respect to size as shown in the statistical distribution of annual plant capacities in Figure 3. A composite model plant was selected as providing the required background for the work conducted for this study. The composite plant consists of melting facilities capable of making 996,600 metric tons (1,100,000 net tons) of steel per year. A general flow sheet of the composite plant is shown in Figure 6.

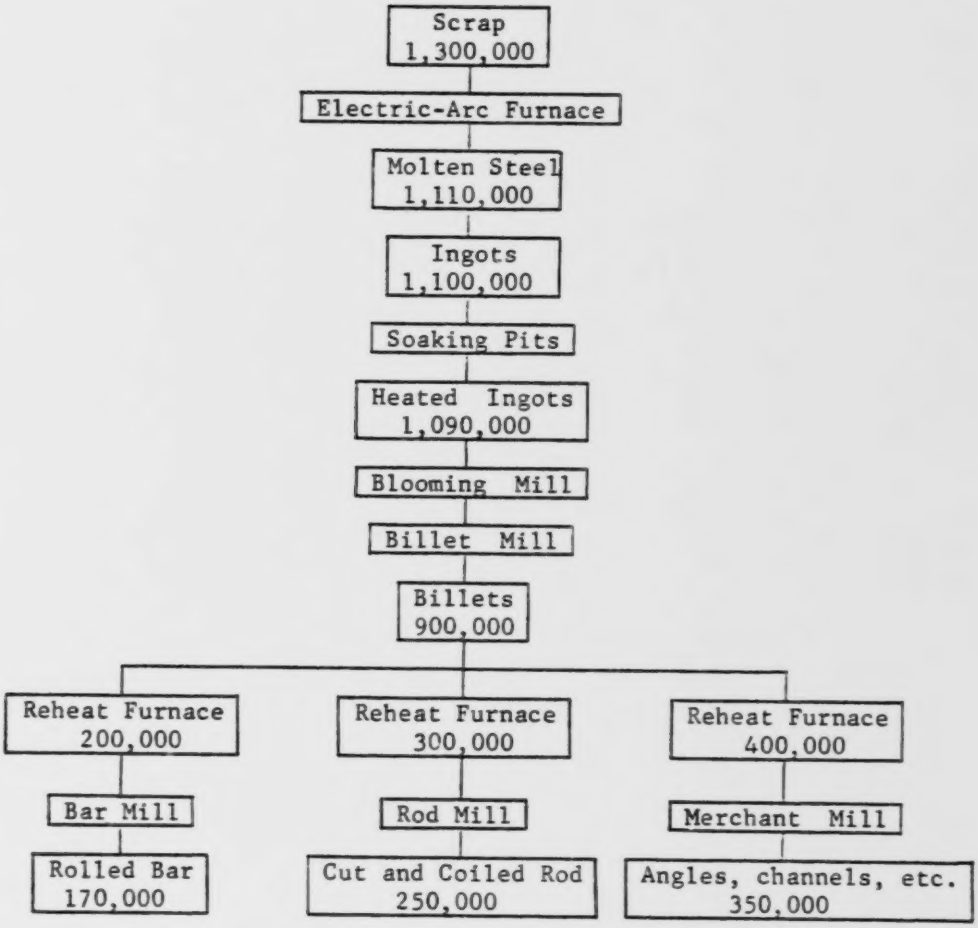

Scrap
1,300,000
Electric-Arc Furnace
Molten Steel
1,110,000
Ingots
1,100,000
Soaking Pits
Heated Ingots
1,090,000
Blooming Mill
Billet Mill
Billets
900,000
Reheat Furnace
200,000
Reheat Furnace
300,000
Reheat Furnace
400,000
Bar Mill
Rod Mill
Merchant Mill
Rolled Bar
170,000
Cut and Coiled Rod
250,000
Angles, channels, etc.
350,000

FIGURE 6. COMPOSITE ELECTRIC-ARC FURNACE PLANT FLOW SHEET (Capacities in net tons)

Electric-arc furnace steelmaking plants placed into operation in recent years almost without exception use continuous casting machines to produce slabs, blooms or billets from the molten steel. For such plants the flow sheet in Figure 6 would not show soaking pits or the blooming mill.

The products, indicated on the flow sheet as bar, rod, angles and channels, would be shipped out of the plant as semi-finished products or used in-plant for the production of special products. The rolling operations all require reheat furnaces to heat the steel to the required rolling temperatures. Fuel requirements for the reheating furnaces are the major consumers of fuel in electric-arc furnace steelmaking plants. Additional fuel requirements are created by reheating furnaces required for forging operations, galvanizing lines, and heat-treating furnaces.

The energy demand of this plant ranges from a minimum of 15.8 x 10^6 MJ/day (15 x 10^9 Btu/day) to a maximum of 24.3 x 10^6 MJ/day (23 x 10^9 Btu/day) with an average energy demand of 21.1 x 10^6 MJ/day (20 x 10^9 Btu/day). Minimum demand usually occurs on weekends and during down time for plant maintenance. Table 11 summarizes the characteristics of the steel plant model.

TABLE 11. STEEL PLANT STATISTICS

	metric ton/year	(ton/year)
Molten steel capacity	996,600	(1,110,000)
Rolled bars	154,000	(170,000)
Cut and coiled rod	226,000	(250,000)
Angles, channels	317,000	(350,000)
Energy Demand	10^6 MJ/day	(10^9 Btu/day)
Maximum	24	(23) (peak production)
Average	21	(20) (normal production)
Minimum	16	(15) (weekends, downtime)

Model Plant Relationship to Industry

The model plant selected for the model is not representative of any particular electric-arc steelmaking plant in the United States. Rather, it is a hypothetical plant which is to be used as a representative of the industry. The model plant would be located on a river where water would be available for the gasification plant and advantage could be taken of river transportation of coal. The majority of the electric-arc furnace plants in the United States are located on rivers or coastal sites where there is a ready access to water.

Gasification Plant Design

A Koppers-Totzek gasification system was selected for the steel plant model. This plant would consist of four two-headed gasifier units and would gasify about 1721 metric ton/day (1900 ton/day) of coal producing approximately 23.5 x 10^6 MJ/day (22.3 x 10^9 Btu/day) of fuel gas with a high heating value of 11.3 MJ/Nm^3 (286 Btu/scf). An MDEA cleanup system was selected for use in desulfurizing the fuel gas. This process, which operates at atmospheric pressure, is commonly recommended by Koppers in such applications. Figure 7 shows an overall flow sheet of the gasifier and gas-cleanup system.

The Koppers-Totzek process is restricted primarily to oxygen-steam gasification producing intermediate-energy gas of about 11.8 MJ/Nm^3 (300 Btu/scf). The intermediate-energy gas was considered desirable for the steel mill application after reviewing the combustion requirements of the various furnaces in the plant. Conversion of these processes to a gas with a lower heating value would require extensive conversion and modification and was considered unfeasible.

The coal selected for the steel mill model was a lignitic coal with a moisture content of about 27 percent and a sulfur content of 1.7 percent. A complete analysis of the coal is given in Table 12. This coal would be pulverized to 70 percent through a 200-mesh screen and dried during the pulverization step to less than 4 percent surface moisture before feeding to the gasifier.

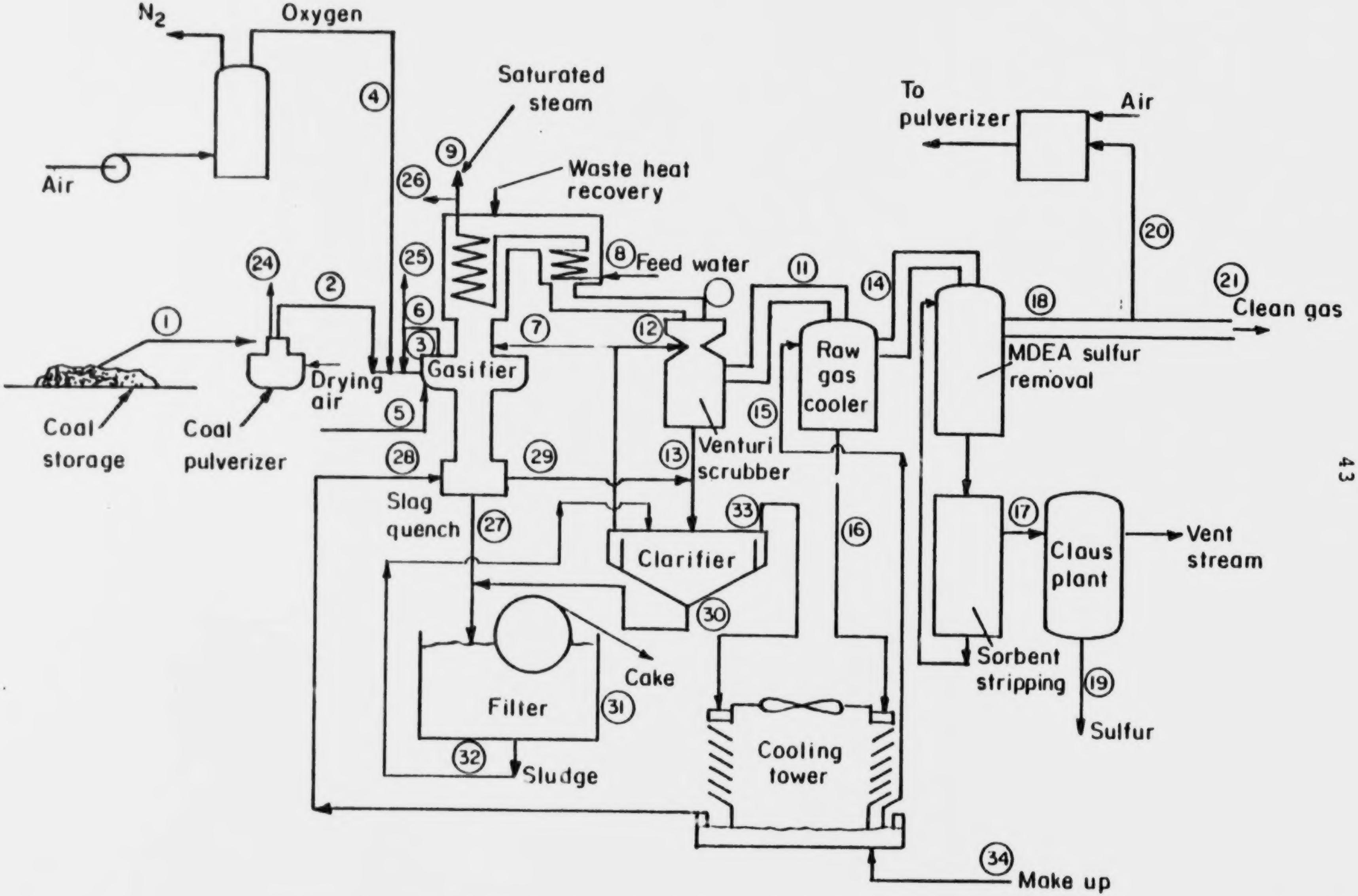

N2
Oxygen
Air
④
Saturated steam
⑨
Waste heat recovery
㉖
To pulverizer
Air
⑳
㉔
②
㉕
⑥
⑧
Feed water
⑪
⑭
㉑
Clean gas
①
③
⑦
⑫
⑱
Gasifier
Raw gas cooler
MDEA sulfur removal
Coal storage
Coal pulverizer
Drying air
⑤
⑮
㉘
㉙
⑬
Venturi scrubber
Slag quench
㉗
㉝
⑯
⑰
Claus plant
Vent stream
Clarifier
㉚
Cake
Filter
㉛
Sorbent stripping
⑲
Sulfur
Cooling tower
㉜
Sludge
㉞
Make up

FIGURE 7. KOPPERS-TOTZEK/MDEA STEEL MILL GASIFICATION PLANT

TABLE 12. COAL ANALYSIS FOR STEEL MILL MODEL

Proximate (as received)	
Moisture .	26.5 percent
Ash	10.1 percent
Volatiles	29.0 percent
Fixed Carbon	34.4 percent
Heat Value	8416 Btu/lb
Ultimate (dry)	
Carbon	66.10 percent
Hydrogen	4.46 percent
Nitrogen	0.67 percent
Chlorides	0.07 percent
Sulfur	1.70 percent
Ash	13.40 percent
Oxygen	13.60 percent
Hardgrove Grindability	52.2
Ash Fusion Temperature (initial atmosphere)	
Initial deformation	1141 C (2087 F)
Softening	1224 C (2236 C)
Fluid	1289 C (2352 F)

Gas produced in the gasification step would be passed through a waste-heat boiler for steam recovery and then through a two-stage venturi scrubber for removal of particulates and any tars that may be formed in the process. Because the Koppers-Totzek gasifier operates at such a high temperature (about 1815 C [330 F]) only trace amounts of tars, phenols, oils, and a relatively small amount of ammonia are present in the fuel gas. From the venturi scrubber the gas would be processed through a cooler and then into the MDEA sulfur-removal system for removal of sulfur compounds.

The final fuel gas would contain about 300 ppm of sulfur compounds and the overall efficiency of the gasification process would be about 70 percent. A complete material balance for the gasification plant is given in Appendix B. Pertinent statistics on the gas plant design are summarized in Table 13.

TABLE 13. GASIFICATION PLANT DESIGN
FOR STEEL MILL MODEL

Gasifier - Koppers-Totzek (four two headed unit trains)

Desulfurization - MDEA (methyl-diethanolamine) with Claus sulfur recovery

Maximum Gas Production Rate - 23.5 x 10^6 MJ/day (22.3 x 10^9 Btu/day)

Minimum Gas Production Rate - 15.3 x 10^6 MJ/day (14.5 x 10^9 Btu/day)
(with all four units operating)

Gas High Heat Value - 10.7 MJ/Nm^3 (286 Btu/scf)

Coal Consumption - 1730 metric ton/day (1900 ton/day)

Overall Efficiency - 67.3 percent

The four-unit gasification plant design was selected to provide adequate flexibility for steel mill operations. The design capacity of 23.5 x 10^6 MJ/day (22.3 x 10^9 Btu/day) is slightly less than the maximum steel plant demand of 24.3 x 10^6 MJ/day (23 x 10^9 Btu/day). The Koppers unit can be operated at up to 10 percent over capacity in situations such as this where peak demands are intermittent and not for sustained periods such as in the case of the steel mill. Therefore the Koppers plant shown would be capable of up to 25.8 x 10^6 MJ/day (24.5 x 10^9 Btu/day) during peak periods.

A complete standby gasifier unit was not felt necessary in this particular case. Steel mills have planned annual outages for maintenance and also low-load periods on weekends and during certain times of the year. The Koppers gasifier has demonstrated a high availability of up to 95 percent in foreign installations and presumably most maintenance could be taken care of during steel mill down times or low-load period. The four-unit design would allow one complete train to be out for service while the plant would still be able to maintain the minimum mill energy demand of 15.8 x 10^6 MJ/day (15 x 10^9 Btu/day). This would allow for gasifier maintenance and repair on weekends and other low-load periods without disrupting service.

Table 14 gives energy flows in the steel mill gasification plant.

TABLE 14. ENERGY BALANCE ON STEEL MILL GASIFICATION PLANT
(Metric and English)

Metric	English
Energy Produced	
Product Gas - 23.5 x 10^6 MJ/day (980 x 10^3 MJ/hr)	22.3 x 10^9 Btu/day (928 x 10^6 Btu/hr)
Saturated Steam	
Waste Heat Boiler - 75,963 Kg/hr (262 C and 4.72 x 10^6 Pa)	167,688 lb/hr (503 F and 685 psi 37 x 10^6 Btu/hr for turbine expansion
Jacket Steam - 14,179 Kg/hr (135 C and 2.07 x 10^5 Pa)	31,300 lb/hr (275 F and 30 psig)
Energy Consumed	
Gasifier Steam - 11,808 Kg/hr	26,032 lb/hr
Oxygen Plant - 52.7 x 10^3 MJ/hr	50 x 10^6 Btu/hr
Sulfur Removal and Recovery(16)	
Low Pressure Steam - 861 Kg/hr	1900 lb/hr
Miscellaneous (pumps, etc.)(16) - 17.6 x 10^3 MJ/hr	16.7 x 10^6 Btu/hr
Net Energy Requirements	
Oxygen Plant - 14.77 x 10^3 MJ/hr	14 x 10^6 Btu/hr
Miscellaneous - 17.6 x 10^3 Mj/hr	16.7 x 10^6 Btu/hr
Total - 32.37 x 10^3 MJ/hr	30.7 x 10^6 Btu/hr
Overall Efficiency	
Product Gas Produced - 979 x 10^3 MJ/hr	928 x 10^6 Btu/hr
Energy Required - 32.27 x 10^3 MJ/hr	30.7 x 10^6 Btu/hr
Net Energy Production - 946.6 x 10^3 MJ/hr	897.3 x 10^6 Btu/hr
Coal Consumed - 71,777 Kg/hr (19.6 MJ/Kg)	158,448 lb/hr (8416 Btu/lb) = 1333 x 10^6 Btu/hr
Overall Efficiency - 946.6/1407 = 0.673	897.3/1333 = 0.673

As mentioned, the plant produces 23.5 x 10^6 MJ/day (22.3 x 10^9 Btu/day) of product gas or 979 x 10^3 MJ/hr (928 x 10^6 Btu/hr) at design capacity after requirements for coal drying are satisfied. In this design saturated steam at 503 F and 685 psig (167,000 lb/hr) is generated in the waste heat boiler* and an additional 14,179 Kg/hr (31,300 lb/hr) of low-pressure steam is generated in the cooling water jacket*.

The oxygen plant is the biggest single energy consumer in the plant and is estimated to require 52.8 x 10^3 MJ/hr (50 x 10^6 Btu/hr) which is based on an oxygen plant energy-consumption rate of 299 kwhr/metric ton (330 kwhr/ton)(14,15). This energy consumption is for compressing air to about 690 kN/m^2 (100 psig) prior to separation into oxygen and nitrogen. The saturated steam from the waste-heat boiler can be used to supply about 39.0 x 10^3 MJ/hr (37 x 10^6 Btu/hr) in a turbine condensor unit leaving 2.4 x 10^3 MJ/hr (9 x 10^6 Btu/hr) still required.

The MDEA sulfur removal process requires steam for regenerating the amine absorbent. Some steam, however, is generated in the Claus sulfur recovery unit partially offsetting the regeneration requirement. The net steam requirement is estimated to be about 861 Kg/hr (1900 lb/hr) of low pressure steam.** This requirement could be satisfied by the 2386 Kg/hr (5268 lb/hr) of jacket steam that remains after the gasifier steam demand of 11,792 Kg/hr (26,032 lb/hr) is satisfied.

Miscellaneous energy requirements for the plant (pumps, fans, etc.) are estimated at about 1.8 percent of the product gas energy production.** This requirement of 27.1 x 10^3 MJ/hr (25.7 x 10^6 Btu/hr) as well as the additional 9.5 x 10^3 MJ/hr (9 x 10^6 Btu/hr) required by the oxygen plant would be satisfied by purchased electricity. Electric arc secondary steel plants are major consumers of electricity and as such would have adequate supply and low rates making this attractive. This would have to be evaluated in detail however before a final decision could be made.

---

*In some designs high pressure steam is generated in the waste-heat boiler. However, this leads to higher tube temperatures and hence higher corrosion rates due to acid gas (H_2S, etc.) attack. In this design maximum tube life was desired and, therefore, steam temperatures were limited to saturation.

**Based on data in Reference (10) for nearly identical plant. Values were adjusted based on plant energy production rate (clean gas).

Figure 8 shows a representative plot plan of the steel mill with the gasifier and coal-storage piles included. The steel mill proper covers about 930,000 m^2 (230 acres) of land with approximately 280,000 cm^2 (70 acres) of land in building area. Though the buildings in Figure 8 are shown linked together, in reality they would be interspersed throughout the available area and relatively little free land would be available in the steel mill proper. The gasification and gas-cleaning plant is estimated to require about 630,000 m^2 (15 acres) of land and coal storage for a 1-month supply of coal would require an additional 10,000 m^2 (2-1/2 acres).

Burners and Furnaces in a Secondary Steel Plant

This section describes the types of burners used to fire various processes in the model steel plant and the possibilities for firing these burners with an intermediate-energy fuel gas produced by coal gasification.

Burner Types

A great portion of the burners, especially those of larger capacity, in the secondary steel industry are designed with provision to burn either coke-oven gas or natural gas. As indicated elsewhere (Appendix A) some of the potential compositions produced by coal gasification systems are within the range of coke-oven gas composition, and several have burning parameters within the coke-oven gas range. Thus, those burners which can be converted with little difficulty from natural gas to coke-oven gas also have a potential for conversion to some intermediate heating value gases. The following figures illustrate several of the burner types used in the secondary steel industry.

Figures 9 and 10 show two designs of flat-flame burners. Swirl is imparted to the air in these burners, so that the flame spreads out along the surface and heats it to a point where there is a significant radiant heat output. Several manufacturers make this type of burner. Such burners might be found in the furnaces in the forged products and wire making department.

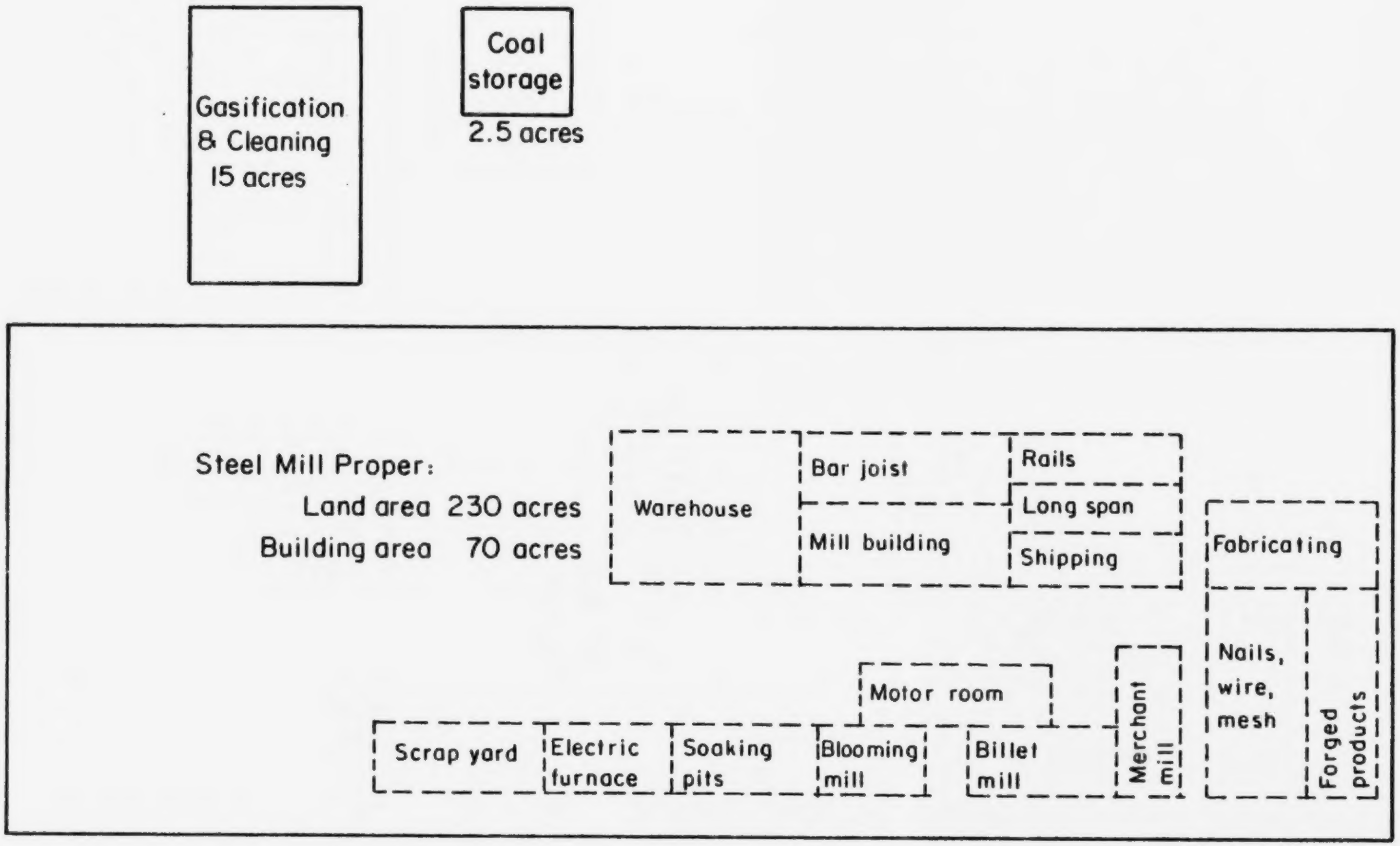

Gasification
& Cleaning
15 acres
Coal
storage
2.5 acres
Steel Mill Proper:
Land area 230 acres
Building area 70 acres
Warehouse
Bar joist
Mill building
Rails
Long span
Shipping
Fabricating
Nails,
wire,
mesh
Forged
products
Motor room
Scrap yard
Electric
furnace
Soaking
pits
Blooming
mill
Billet
mill
Merchant
mill

FIGURE 8. STEEL MILL PLOT PLAN

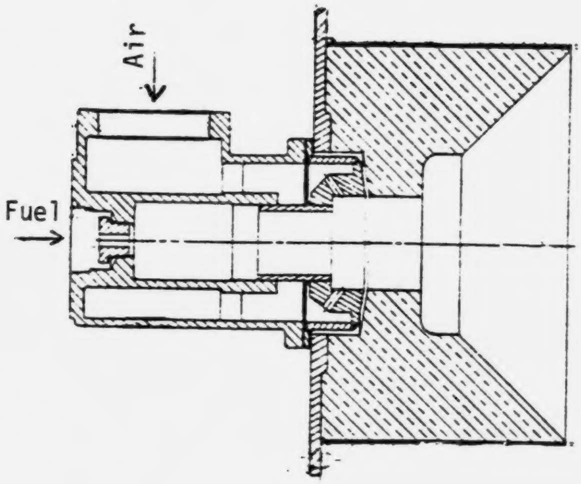


FIGURE 9. BLOOM HTR FLAT-FLAME NOZZLE-MIX BURNER

The burner is designed to use natural gas or coke-oven gas in a sealed-in tile.

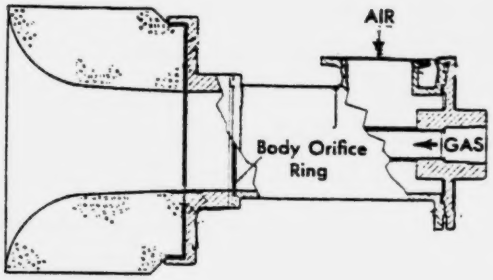

AIR
Body Orifice
Ring
GAS

FIGURE 10. NORTH AMERICAN 4832 FLAT-FLAME (OR RADIATION TYPE) NOZZLE-MIX BURNER

The burner is designed to use natural gas or coke-oven gas in a sealed-in tile.

These burners are designed so that they can successfully fire coke-oven gas. However, the maximum spud size may not be sufficient to obtain an acceptable fuel velocity of an intermediate heating value gas (changing from 17.7 MJ/Nm^3 [450 Btu/ft^3] to 11.8 MJ/Nm^3 [300 Btu/ft^3] fuel gas requires a 50 percent increase in volume flow rate). Furthermore, the intensity of swirl imparted to the fuel by the combustion air might be insufficient to obtain satisfactory combustion. Thus, even though the stability parameters might match those of coke-oven gas, the burner may not perform satisfactorily and could need replacing. A short series of experiments on a few of these burners would be the optimum way of answering the interchangeability question.

Figure 11 shows one version of a forced-air radiant-tube burner. Flame holder details vary with manufacturer. Other designs of burner, including inspirators (in which the fuel aspirates the air) and exhaust suction type are available. These burners were used in the forged products department, and for annealing. For burners such as in Figure 11, there appears to be no reason that intermediate fuels could not also be used. Possibly, a change in spreader might be needed, and, definitely, a somewhat increased back pressure on the fuel would be required. For systems using inspirators, the chances are that the inspirator portion would have to be changed.

Figure 12 shows a general heating burner that can use a variety of fuels, and can be fired with considerable excess fuel. It is used in various operations such as primarily in the forged-products area. This burner should operate satisfactorily on intermediate-heating value gas, but the back pressure would have to be increased to 16 percent to 24 percent of the air pressure.

Figure 13 shows a dual-fuel ultrastable burner only used in limited numbers in forging operations. From the pressure requirements on this burner for coke-oven gas (see Figure 13), it appears quite questionable that it could be used for intermediate-heating value fuels. A change in burner design or burner type would probably have to be made.

Figure 14 shows a burner that can produce a long flame or operate with high excess air. It is used in forging operation and wire making. From the low pressure drop with coke-oven gas in this burner and the great flexibility of operation, there should be no difficulty in utilizing the intermediate-heating value fuel from the gasifier in the burner.

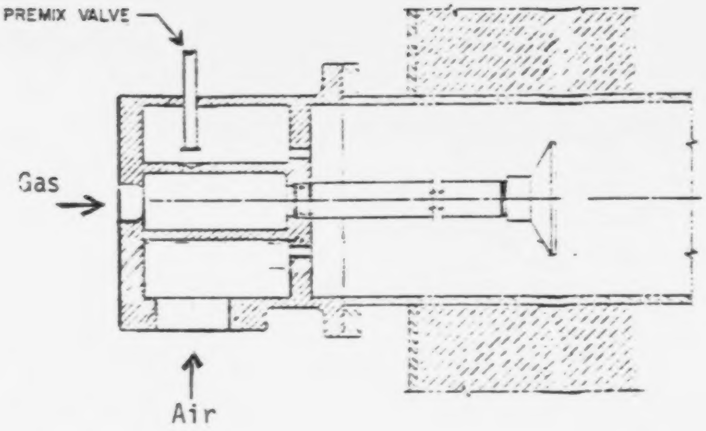

PREMIX VALVE
Gas
Air

FIGURE 11. BLOOM FORCED-AIR RADIANT TUBE BURNER

Develops a Long Flame With Uniform Heat Release Along the Length of a Radiant Tube. Natural Gas or Coke Oven Gas May Be Used.

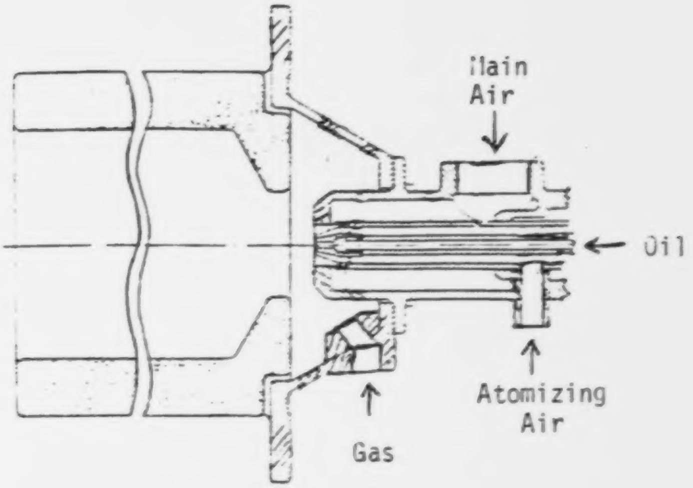


FIGURE 12. NORTH AMERICAN 220 AND NORTH AMERICAN 221 DUAL-FUEL NOZZLE-MIX BURNERS

General heating burner uses liquified petroleum gas (1 percent of air pressure), natural gas (2 percent of air pressure), coke-oven gas (8 percent of air pressure), and light oil in sealed tile. Burner will operate at "double-rich" condition.

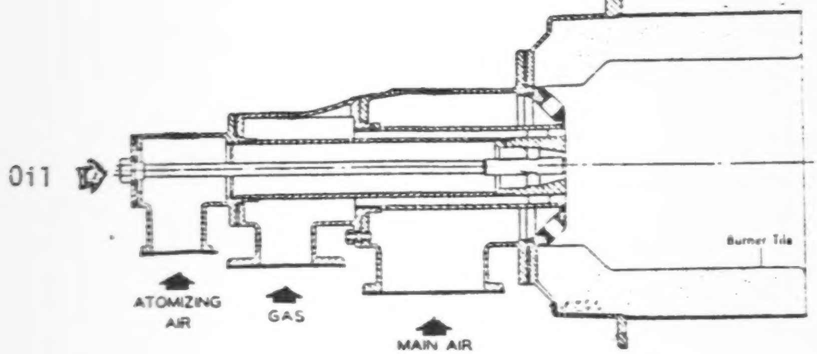

Oil
ATOMIZING AIR
GAS
MAIN AIR
Burner Tile

FIGURE 13. NORTH AMERICAN 214 DUAL-FUEL NOZZLE MIX BURNER

Ultrastable burner uses natural gas, coke-oven gas, light oil, or heavy oil in sealed-in tile. For coke-oven gas (19.7 MJ/Nm^3) (500 Btu/scf, 0.4 sp.gr), gas pressure equals 1/5 air pressure at stoichiometric; for "double-rich" firing, coke-oven gas pressure must equal air pressure.

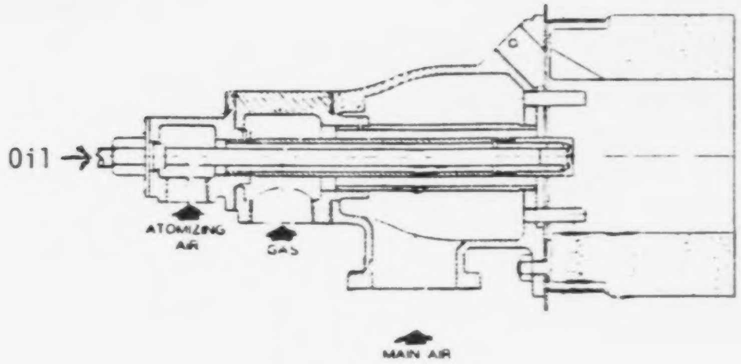

Oil
ATOMIZING AIR
GAS
MAIN AIR

FIGURE 14. NORTH AMERICAN 223 DUAL-FUEL NOZZLE-MIX BURNER

Excess air burner uses natural gas, coke-oven gas (413 Pa gage) (0.06 psia) or distillate oil in a sealed-in tile. Mixture rate varies from 50 percent excess air, the amount depending on burner size and firing rate.

Figure 15 shows a small long-flame burner used on the rolling mills in conjunction with larger long-flame burners such as shown in Figure 16. Burners similar to that shown in Figure 15 are also used in the soaking pits and rod mills. It appears probable that both burners could be used with intermediate heat-up value fuel with little difficulty.

Figure 17 shows a premix radiant cup burner of relative small capacity, often used in large numbers for annealing and similar operations. The burner should be quite adaptable to intermediate-heating value fuels. While tips are not listed as available for 11.8 MJ/Nm^3 (300 Btu/scf) fuels. demand for such should lead to their production. However, the mixing and monitoring system would probably also require some revision.

Figure 18 is a typical ring-type gas and oil burner for boilers. This particular burner can be stretched to handle gases down to 15.7 MJ/Nm^3 (400 Btu/scf). However, below this value, down to 9.8 MJ/Nm^3 (150 Btu/scf), a different burner design would be recommended. Below 9.8 MJ/Nm^3 (250 Btu/scf), a third burner design would be used.

Summary of Burner Changes

The analysis of the study of the secondary steel plant is based on the assumption that a Koppers-Totzek gas is used as a replacement for natural gas. On this basis, it is found that most of the burners used are in the questionable area as to satisfactory performance. Most of the burners are built to handle natural gas and coke-oven gas. The stability values for Koppers-Totzek gas are better than natural gas (except for premixed burners where flashback may occur), and about equal to coke-oven gas. However, the lower heating value (order of 11.8 MJ/Nm^3 [300 Btu/scf]) compared with coke-oven gas (order of 19.7 MJ/Nm^3 [500 Btu/scf] can lead to flow distortions in nozzle mix burners (because of the higher volume fuel flow rate). This could result in unsatisfactory performance. Experimental data are required on some typical industrial burners using Koppers-Totzek fuel to answer this question in a definitive manner. The small numbers of premix burners must be considered carefully, first as to changes needed to prevent flashback, and, second, as to changes needed in the mixing system; new burners, or a completely different

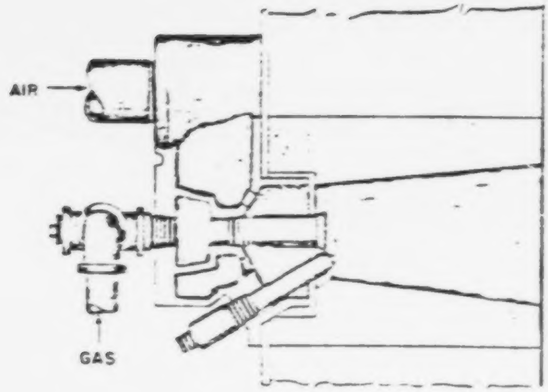


FIGURE 15. BLOOM 401-L LONG-FLAME BURNER

Natural gas, propane, butane, or coke-oven gas. Turndown ratio is 20:1. At rated capacity 0.2 psi required for both gas and air.

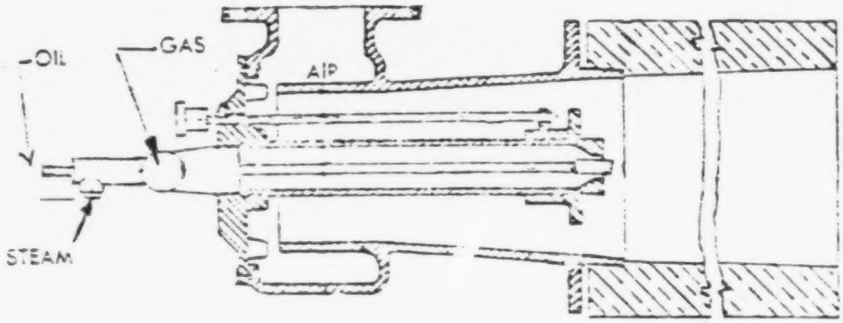


FIGURE 16. BLOOM LONG-FLAME BURNER, COLD AIR

Radial fins result in uniform air flow. When the adjustible flange in in the back position, the flame in luminous. With the flange in the forward position, the flame has little luminosity.

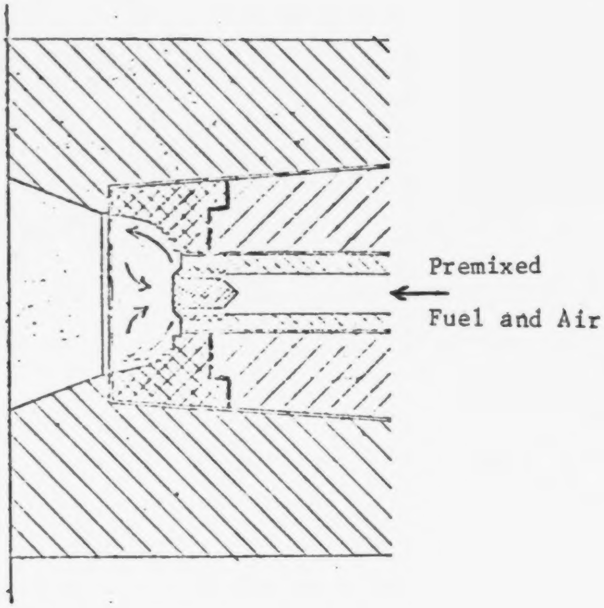

Premixed
Fuel and Air

FIGURE 17. SELAS DURADIANT PREMIX BURNER

Cup-shaped ceramic "washed" by hot combustion products, radiates heat to work. Tips are available for gases with 15.8 to 126.1 MJ/Nm^3 (400 to 3200 Btu/scf).

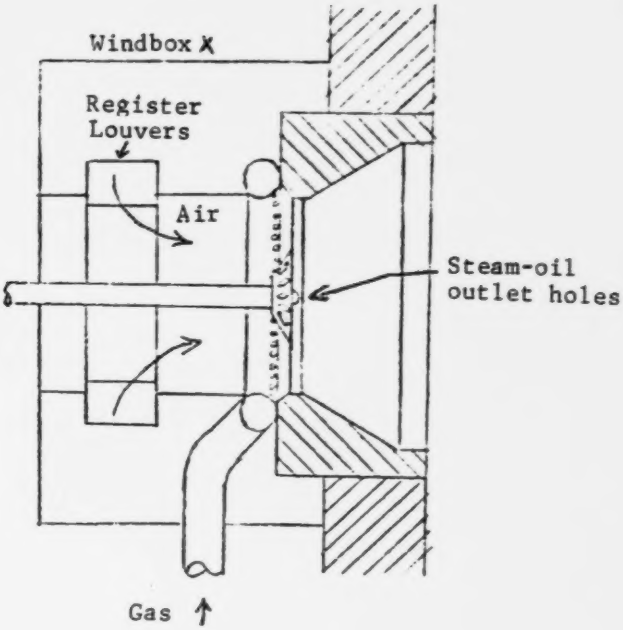

Windbox
Register Louvers
Air
Steam-oil outlet holes
Gas

FIGURE 18. ERIE CITY RING-TYPE GAS AND OIL BURNER FOR BOILER USE

This burner will handle gases with heating valued down to 15.8 MJ/Nm^3 (400 Btu/scf) with some adjustments.

design might well be needed. This is especially true in the case of space heaters. In some instances, boiler burners may be easily adaptable to Koppers-Totzek gas, but in others, it is expected that new burners will be required.

Using Koppers-Totzek gas, no difficulty is foreseen relative to radiation output changes or furnace pressure drops. Fuel line pressure drops, on the other hand, require careful consideration.

If the heating value of an intermediate-energy gas could be increased to a value similar to that for coke-oven gas (either by removal of inerts or addition of a higher grade fuel such as methane) conversion of burners in a secondary steel plant could be greatly simplified.

IV. CONVERSION OF A REFINERY TO LOW-ENERGY GAS

The Refinery Industry

In order to generalize the conclusions for the hypothetical refinery (to be described later), it is necessary to examine how this refinery compares with other refineries in the United States. The following discussion presents data on U.S. refineries so that the model refinery can be compared to other refineries. Among items discussed are

- Energy consumption
- Types of fuels used
- Amenability to conversion to low- or intermediate-energy gas.

Energy Consumption

Size. The most important variable affecting the total energy consumption of a refinery is the size of the refinery, expressed in terms of crude oil throughput. The distribution of sizes of the refineries in the United States is shown in Table 15. The median capacity is 4.53×10^6 liter/day (28,500 barrel/day). That is, half the refineries in the United States are smaller than this and half are larger. The average capacity is 9.11×10^6 liter/day (57,318 barrel/day), this being higher than the median because of a relatively few numbers of very large refineries.

Complexity. Another secondary variable affecting the energy consumption of a refinery is the complexity of the processing operations used. W. L. Nelson(18) has quantified the complexity of refineries by defining a parameter known as the Nelson complexity factor. This factor is obtained by multiplying the capacity of each type of processing (distillation, catalytic cracking, etc.) by a factor, then summing the products, and then dividing the total by the crude oil capacity. The complexity factors for the various refining processes are shown in Table 16. Also shown in this table are some approximate energy requirements for the processes. Although the complexity factors were originally based on costs, they have been found to be reasonably good

TABLE 15. U.S. REFINERY SIZE DISTRIBUTION AS OF JANUARY 1, 1975(17)

Capacity Range, 10^3 B/CD(a)	Number of Refineries		Total Capacity, B/CD(a)	Percent of Capacity	Average Capacity, B/CD(a)
<5	49	124	146,592	0.99	2,992
5-10	31		230,688	1.55	7,442
10-15	19		234,780	1.58	12,357
15-25	25		517,520	3.49	20,701
25-50	50		1,910,592	12.87	38,712
50-75	21		1,309,385	8.82	62,352
75-100	21		1,878,950	12.66	89,474
100-200	28		4,002,900	26.96	142,961
>200	15		4,614,000	31.08	307,600
TOTAL	259		14,845,407	100.00	57,318

Median Capacity (128 refineries smaller, 128 refineries larger) = 28,500 B/CD

(a) B/CD = barrels per calendar day. 1 barrel = 42 gallons = 158.97 liters.

measures of the unit energy consumption (energy consumption per unit of throughput) of the various processes. The Nelson complexity factor for the U.S. refinery industry as a whole is 8.88. This value is based on the total capacities of the various processes in the U.S., and as such it represents an "average" U.S. refinery (not a median size refiner).

Nelson has developed correlations of the energy consumptions of refineries in terms of the refinery complexity and the fuel cost. The latter is important because, as fuel costs have risen, more extensive conservation measures have been adopted with the result that energy consumption has decreased. This is shown in Figure 19, which is a plot of energy consumption versus fuel cost with complexity as a parameter. This plot covers the time period from 1950 through 1975. The years corresponding to the various points (fuel costs) are shown for the center curve. Figure 20 shows the energy consumption as a function of complexity with fuel cost (time) as a parameter. The unit energy consumption varies linearly with complexity.

TABLE 16. COMPLEXITY FACTORS AND ENERGY REQUIREMENTS FOR REFINING PROCESSES(18)

	Nelson Complexity Factor	Approx. Requirements Per Barrel of Feed		
Refining Operation		Fuel, MJ	(10^3 Btu)	Steam, lb
Topping, low gravity crude, light duty	1	55	(52)	40
Topping, high gravity crude, heavy dury	1	74	(70)	55
Vacuum flash	1			
Vacuum distillation	2			
Thermal Cracking				
Two coil	3	506	(480)	45
Visbreaking	2	200	(190)	30
Catalytic crude oil		422	(400)	40
Catalytic Cracking				
65 percent conversion (zero recycle)	5			
Airlift type		501	(475)	40
Fluid bed		480	(455)	50
75 percent conversion (0.75 recycle)	6			
Airlift type		601	(570)	75
Fluid bed		591	(560)	90
Thermal reforming	3	174	(165)	20
Catalytic reforming	4	338	(320)	75
Alkylation(a)	10	12.7	(12)	680
Polymerization(a)	9	200	(190)	75
Isomerization	3			
Hydrogen treating, 100 ft^3/bbl	2			
Hydrogen treating, 300 ft^3/bbl	4			
Hydrocracking	6			
Lube manufacture				
Solvent extraction	4.5	150	(142)	130
Solvent dewaxing	9	--	--	300
Lube finishing	50	166	(157)	100
Wax finishing	100	111	(105)	300
Typical complete lube plant	62			
Coking	5			
Asphalt manufacture	2			

(a) Fuel and steam requirements per barrel of product (rather than feed).

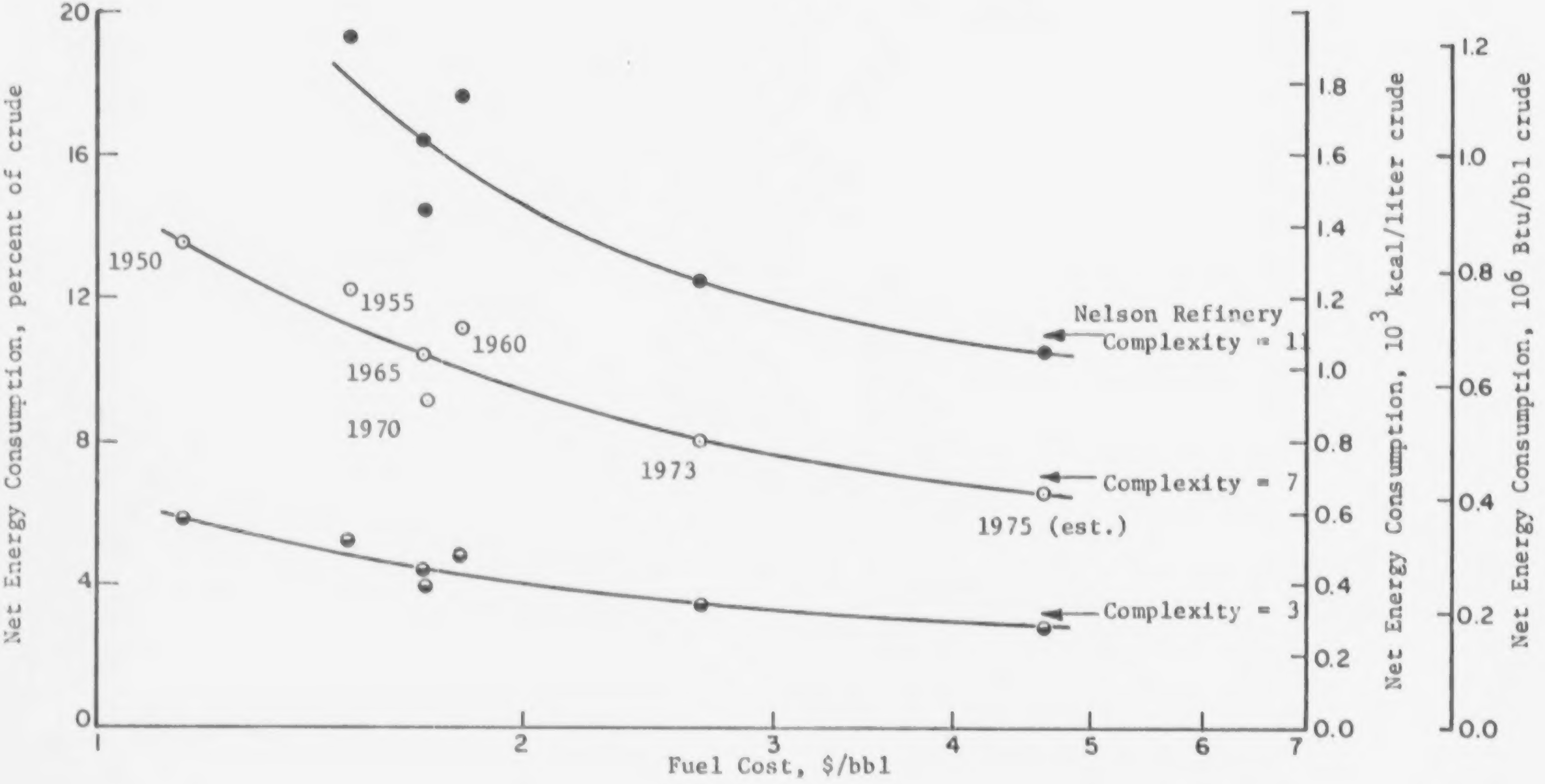


FIGURE 19. REFINERY ENERGY CONSUMPTION VERSUS FUEL COST(19)

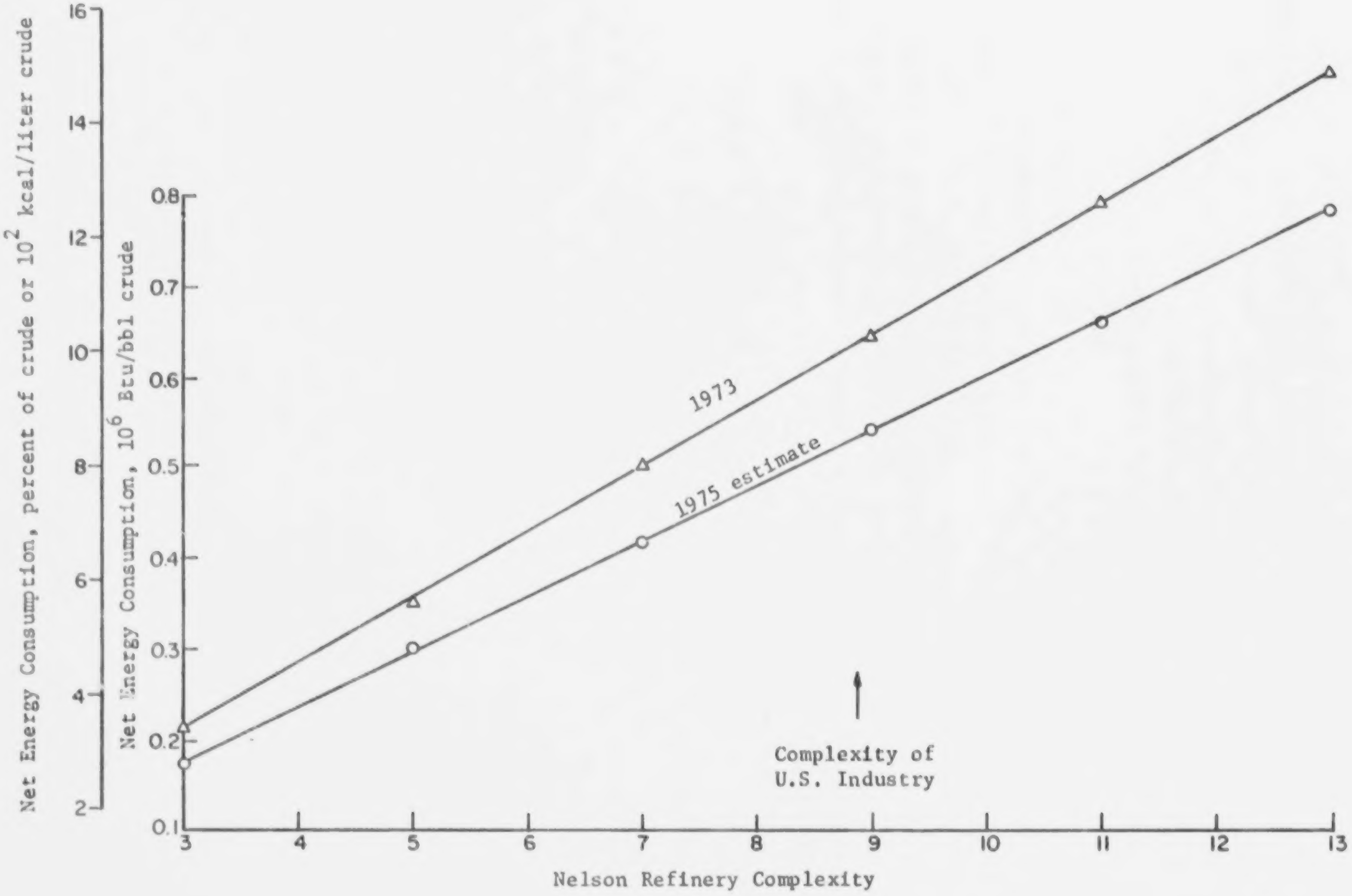


FIGURE 20. REFINERY ENERGY CONSUMPTION VERSUS REFINERY COMPLEXITY(19)

Refinery Energy Consumption Data. Knowing that the size variable can be accounted for by using unit energy consumptions (energy consumption per unit of throughput) and having Nelson's correlations for the complexity variable, typical refinery energy consumption data are needed to compare the energy consumption of the model refinery with that of the U.S. refining industry as a whole. The Bureau of Mines publishes data on the energy consumed at refineries in the U.S., breaking this down by states (or state groups) and sources of energy(20). Table 17 summarizes the national totals, including the breakdown by sources, for the last 3 years for which the data are available. Table 18 presents the state-by-state breakdown of the energy consumptions for the most recent year available (1973). This table also includes the 1973 and 1975 crude oil capacities for all the states and the unit energy consumptions and average refinery complexities for the states with the larger refinery capacities. As backup information, Table 19 contains the fuel energy contents used by the Bureau of Mines(22) in developing their tabulations.

Types of Fuels Used

Because refinery gas is a major source of energy in most refineries, it will be instructive to consider the quantities of refinery gas available for fuel users in various refineries. Two aspects in which they have available for fuel use are concerned with the two processes which generate the greatest share of the refinery gas at the model refinery--catalytic reforming and catalytic cracking.

Hydrogen From Catalytic Reforming. As mentioned previously, the catalytic reforming process generates as a byproduct considerable quantities of hydrogen. This hydrogen can be used in hydrotreating processes to remove sulfur from liquid fuels. The feed to the catalytic reformer itself requires a mild hydrotreating to remove traces of sulfur which would otherwise poison the catalyst. The quantity of hydrogen needed for other hydrotreating operations depends primarily upon the sulfur content of the crude oil being processed. Also, some refineries use hydrocracking as a conversion process, either instead of, or in addition to, catalytic cracking, and hydrocracking requires relatively large quantitites of hydrogen. There are some refineries at which the

TABLE 17. CRUDE RUNS AND ENERGY CONSUMPTION DATA FOR U. S. REFINERIES(21)

	1971	1972	1973	1974
Crude Oil Capacity, 10^3 B/CD(a)	12,884.31	13,235.09	13,799.62	14,530.85
Crude Run, 10^3 B/CD	11,199.48	11,728.39	12,430.83	12,689.32
Capacity Utilization, percent	86.92	88.62	90.08	87.33
Consumption of Energy Sources				
Oil, 10^3 B	38,072	44,324	49,574	
Liquefied petroleum gas, 10^3 B	6,850	13,418	10,136	
Natural gas, 10^6 scf	1,062,938	1,040,746	1,073,742	
Refinery gas, 10^6 scf	981,557	1,053,492	1,083,363	
Petroleum coke, 10^3 tons	10,444	11,230	13,282	
Coal, 10^3 tons	405	339	329	
Electricity, 10^6 kwhr	20,720	22,612	23,382	
Steam, 10^6 lb	36,762	38,870	33,945	
Energy Consumption, 10^9 Btu				
Oil	239,359	276,318	309,095	
Liquefied petroleum gas	27,475	53,820	40,655	
Natural gas	1,095,889	1,073,009	1,107,028	
Refinery gas	971,742	1,042,957	1,072,529	
Petroleum coke	314,573	338,247	400,054	
Coal	9,728	8,143	7,902	
Electricity	70,697	77,152	79,779	
Steam	44,114	40,644	40,734	
TOTAL	2,773,577	2,910,290	3,057,776	
Energy Consumption, 10^3 Btu/B crude				
Total	678.5	679.8	673.9	
Ex. refinery gas and coke	363.8	357.2	349.4	
Natural gas and LPG only	274.8	263.2	252.9	

(a) Average of values at beginning and end of year. Oil and Gas Journal.

TABLE 18. STATE-BY-STATE BREAKDOWN OF 1973 CRUDE RUNS AND ENERGY CONSUMPTION FOR U.S. REFINERIES(20)

States	1973 Crude Run(e) (10^3 B/CD)	1973 Energy Consumption (10^9 Btu)(e)							1973 Energy Consumption (10^3 Btu/B Crude Run)			Nelson Complexity Factor 1973
		Oil	LPG	Nat. Gas	Ref. Gas	Coke & Coal	Elec. & Steam	Total	Total	Ex. Ref. Gas and Coke	N.G., LPG and Oil	
Arkansas	48.79	1,978	205	5,236	2,914	--	239	10,572				
Calif., Wash., Ore., Alasks, Hawaii	1,971.42	37,942	13,649	146,111	184,618	58,284	25,604	466,748	648.7	310.3	274.8	9.26
Colorado	38.84	1,041	421	2,169	3,086	994	300	8,038				
Delaware, Mass., R.I., Virginia	113.40	4,267	40	1,041	24,744	13,614	8,876	52,582				
Georgia, N. Car., S. Car., Florida	57.76	1,362	373	801	--	--	20	2,555				
Illinois	1,031.12	54,274	2,403	11,182	88,626	42,108	7,363	205,955	547.2	199.9	180.3	8.89
Indiana	491.61	36,944	325	4,542	42,301	15,753	983	100,848	562.0	238.5	233.0	8.11
Kansas	373.27	4,602	999	34,889	30,425	14,126	2,289	87,330				
Kentucky, Tennessee	175.15	9,537	1,656	5,791	12,072	4,790	1,242	35,268				
Louisiana	1,462.09	15,443	6,811	116,835	116,917	32,951	28,060	316,017	592.2	313.2	260.6	9.05
Maryland	18.58	2,223	--	--	11	--	31	2,265				
Michigan	122.88	3,628	650	3,244	7,172	2,319	778	17,971				
Minnesota, Wisc., N. Dak., S. Dak.	242.35	10,575	148	1,524	15,997	9,940	1,501	39,685				
Mississippi, Alabama	287.59	1,776	365	25,411	20,099	4,458	2,975	55,084				
Missouri, Nebraska	101.07	346	2,254	4,446	8,606	4,217	157	20,026				
Montana	119.08	5,042	453	5,653	11,370	5,693	972	29,183				
New Jersey	593.21	38,493	192	9,826	37,280	19,879	10,092	115,763	534.6	270.7	224.0	9.02
New Mexico	46.39	163	610	3,600	3,126	1,175	188	8,862				
New York	100.49	3,911	2,238	--	8,460	2,530	652	17,791				
Ohio	500.32	10,104	734	20,677	47,974	14,456	3,975	97,920	536.2	196.3	172.6	8.52
Oklahoma	447.16	1,374	433	49,739	42,867	13,012	2,351	109,776	672.6	330.2	315.8	9.51
Pennsylvania	604.28	45,763	16	20,390	64,549	21,688	4,481	156,887	711.3	352.0	300.0	10.07
Texas	3,209.11	8,681	4,966	612,382	280,330	113,523	16,033	1,035,915	884.4	548.2	534.5	9.36
Utash	115.97	3,805	586	5,625	8,864	4,789	624	23,933				
West Virginia	13.72	2,534	--	1,235	967	552	263	5,551				
Wyoming	141.23	3,288	128	15,012	10,154	6,385	464	35,431				
Arizona	3.95											
TOTAL	12,430.83	309,095	40,655	1,107,028	1,072,529	407,956	120,513	3,057,776	673.9	349.4	321.1	9.24
Percent of Energy Consumption		10.1	1.3	36.3	35.1	13.3	3.9	100.0				

TABLE 19. FUEL ENERGY CONTENTS USED BY BUREAU OF MINES(22)

Fuel	Energy Content	
	Metric Units	English Units
Crude oil	9,032 kcal/liter	5.675×10^6 Btu/bbl
Distillate fuel oil	9,270 kcal/liter	5.825×10^6 Btu/bbl
Residual fuel oil	10,006 kcal/liter	6.287×10^6 Btu/bbl
Liquefied petroleum gas	6,383 kcal/liter	4.011×10^6 Btu/bbl
Natural gas	9,211 kcal/std m^3	1,031 Btu/scf
Refinery gas	8,845 kcal/std m^3	990 Btu/scf
Petroleum coke	8,400 kcal/kg	30.12×10^6 Btu/ton
Coal	6,699 kcal/kg	24.02×10^6 Btu/ton
Electricity	863 kcal/kwhr	3,412 Btu/kwhr
Steam	669 kcal/kg	1,200 Btu/lb

demand for hydrogen is too great to be satisfied by the reformer byproduct and a separate hydrogen generation plant is used.

The number of refineries in the U.S. using various hydrotreating processes is shown in Table 20. In this table, "hydrotreating" refers to the mildest type of process (such as that used for the reformer feed), "hydrorefining" to more severe processes, and "hydrocracking" to the conversion process mentioned above. The use of hydrotreating processes in this country is expected to increase as more crude oil having higher sulfur content is processed and as the restrictions on the sulfur contents of fuels are tightened.

TABLE 20. NUMBER OF REFINERIES USING HYDROTREATING PROCESSES(17)

Crude Oil Capacity Range, 10^3 B/CD	25	25-100	100	All Sizes
Number of Refineries Using				
Hydrocracking	5	13	26	44
Hydrorefining	2	24	18	44
Hydrotreating	41	82(a)	43	166
All Refineries in U.S.	124	92	43	259

(a) Model refinery included here.

Off Gas From Catalytic Cracking. The catalytic cracking process produces considerable quantities of light hydrocarbons (C_1-C_4) which are collected as a gaseous stream. Much of this stream is made up of unsaturated (olefinic) hydrocarbons such as ethylene, propylene, and butylene. Since these olefins are not as desirable in fuel products as are other types of hydrocarbons, it is common practice to include with a catalytic cracker another process to utilize the olefins produced by the catalytic cracker. The two processes which can be used for this purpose are alkylation and polymerization. Both processes yield a high octane product containing mostly branched-chain paraffin compounds, a product which is blended into gasoline. In alkylation, isobutane is added to the olefins to form branched-chain compounds in the gasoline boiling range. In polymerization, the light olefins combine with each other to form a similar product. A refinery which has either of these olefin utilization processes will not have large quantities of light olefins available for fuel use.

Adaptability to Firing Low-Energy Gas

Land Area. The amount of land area occupied by the processing equipment is important in analyzing the possibility of retrofitting refineries to low-energy gas because it determines the distances over which the gas must be piped. The processing equipment usually occupies only a small part of the total refinery area. Storage tanks usually occupy the largest part of the area. With the increasing emphasis on pollution abatement, water treatment facilities can use a considerable fraction of the refinery area. As examples of typical refinery layout, the plot plans of two refineries recently built in the United States are shown in Figures 21 and 22.

The land area required by a refinery depends on the size and complexity of the refinery. W. L. Nelson(25) has determined some average land usages per unit of refinery throughput and has expressed them in terms of the refinery complexity factor. These data are plotted in Figure 23. They include the land in use for process equipment and storage but not for administration buildings and buffer zones around the plant. Based on some available refinery plot plans, such as those shown in Figures 21 and 22, it appears that the process equipment typically occupies 1/3 to 1/5 of the area included in the correlation of Figure 23.

Access to Waterways. Although not absolutely essential, access to a waterway is an attractive feature of a site for a coal gasification facility. Considerable quantities of cooling water are required for such a facility, even when a recirculating system is used. If the waterway is navigable, it may be desirable to transport at least part of the coal to the facility by barge.

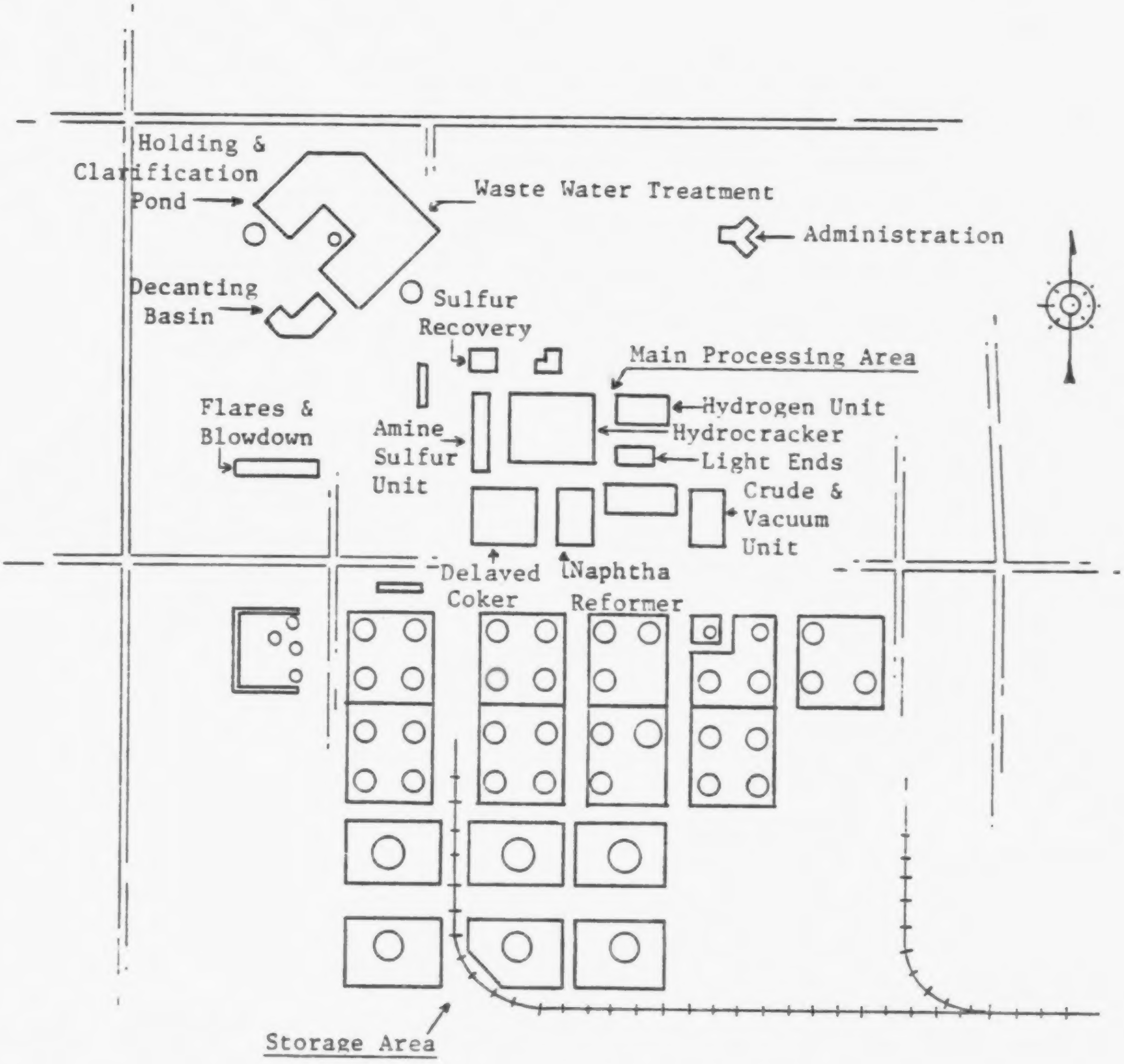

Holding &
Clarification
Pond
Waste Water Treatment
Administration
Decanting
Basin
Sulfur
Recovery
Main Processing Area
Hydrogen Unit
Hydrocracker
Light Ends
Flares &
Blowdown
Amine
Sulfur
Unit
Crude &
Vacuum
Unit
Delayed
Coker
Naphtha
Reformer
Storage Area

FIGURE 21. PLOT PLAN OF ARCO'S CHERRY POINT REFINERY (23)

Capacity: 1,000,000 B/SD
Total Area: 450 Acres

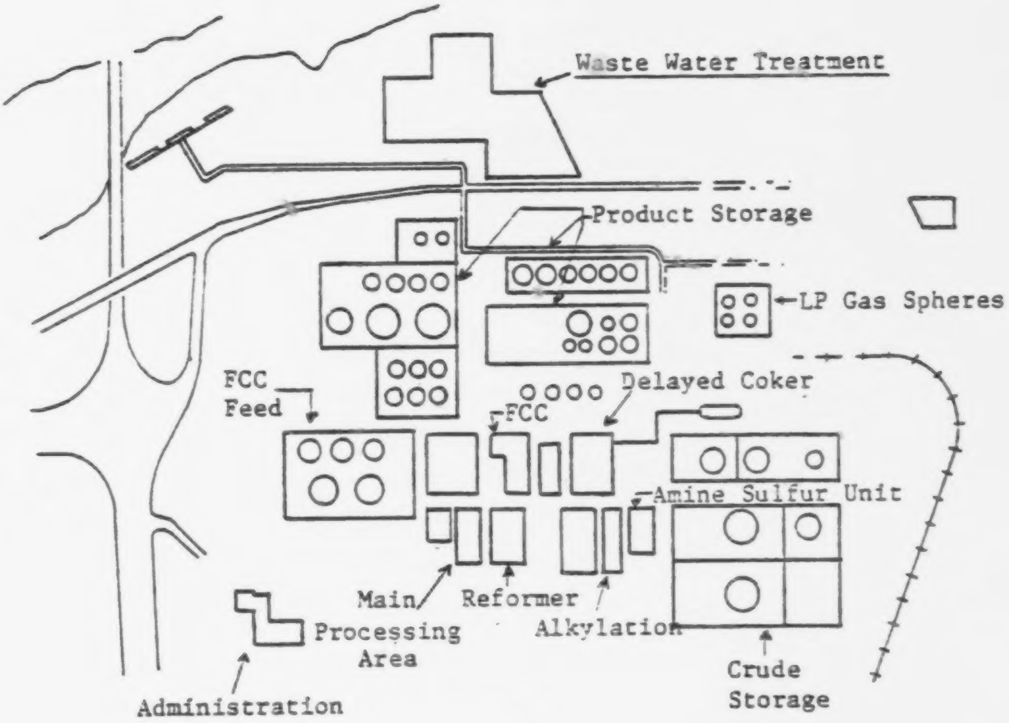

Waste Water Treatment
Product Storage
LP Gas Spheres
FCC Feed
FCC
Delayed Coker
Amine Sulfur Unit
Main Processing Area
Reformer
Alkylation
Crude Storage
Administration

FIGURE 22. PLOT PLAN OF MOBIL OIL'S JOLIET, ILLINOIS, REFINERY(24)
Capacity: 164,000 B/SD

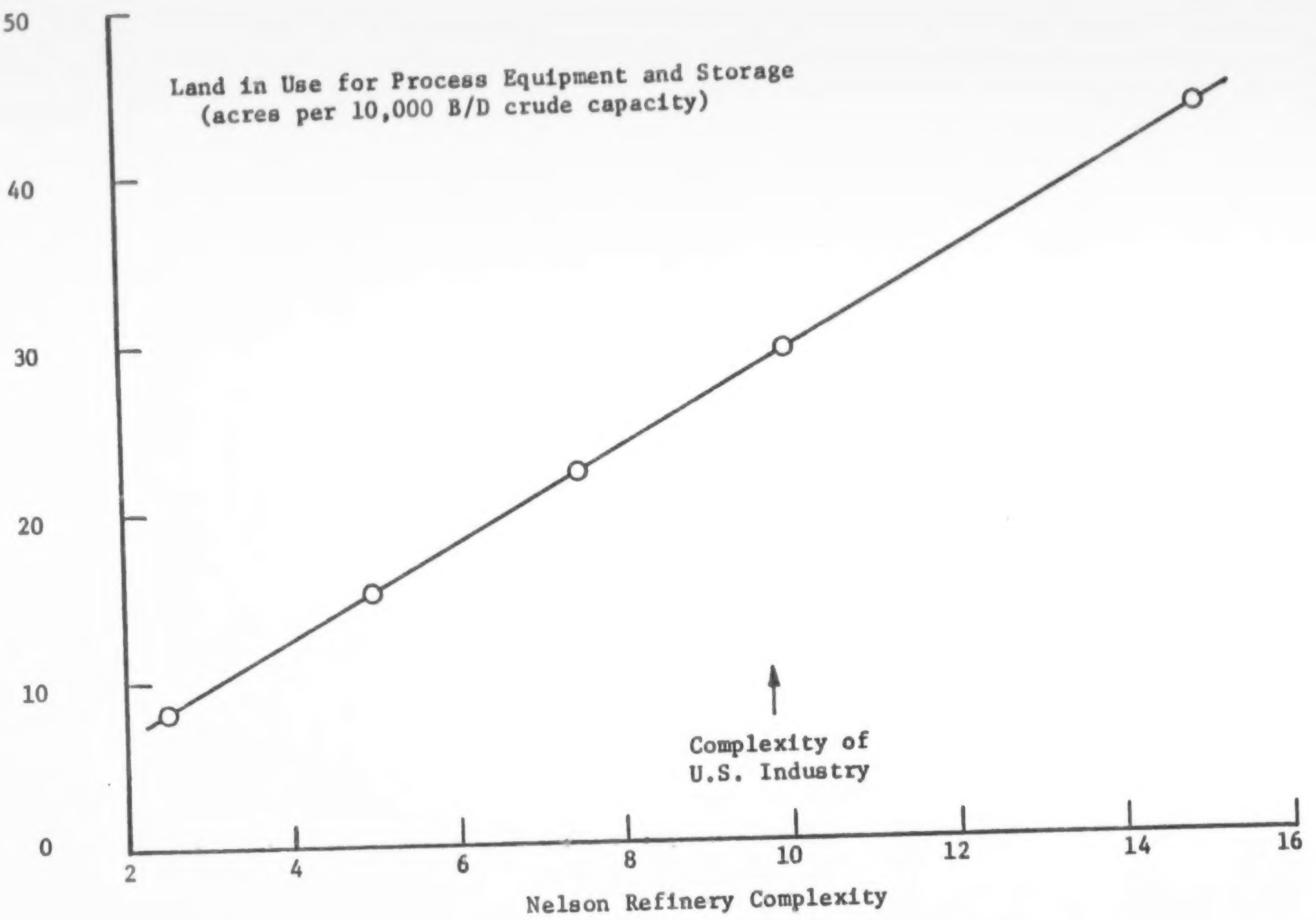


FIGURE 23. LAND IN USE FOR PROCESS EQUIPMENT AND STORAGE AT REFINERIES(25)

Description of Model Refinery

Size and Products

The model petroleum refinery used in this study has a crude oil capacity of about 3.97 x 10^6 liter/day (25,000 barrel/day). The products of the refinery are propane, butane, gasoline, kerosene, distillate, residual (No. 6) fuel oil, and asphalt. There are seasonal variations in the quantities of these products produced. More gasoline is produced in the summer, and more residual fuel oil is produced in the winter. Asphalt is produced only in the summer. Such seasonal variations are normal for petroleum refineries.

Processes

The following refining processes are used in the model refinery:

- Fractionation of crude oil and petroleum fractions
- Catalytic cracking
- Catalytic reforming (including feed hydrotreating)
- Polymerization.

Catalytic cracking is a process for reducing the molecular weight of hydrocarbons and is used to produce hydrocarbons boiling in the gasoline range from higher boiling hydrocarbons. Catalytic reforming and polymerization are processes for producing high octane streams for blending into gasoline. In catalytic reforming, paraffinic and naphthenic hydrocarbons are converted into aromatic hydrocarbons, which high higher octanes. Hydrogen is liberated in this process. In polymerization, light olefins such as ethylene, propylene, and butylenes are combined to form branched-chain hydrocarbons in the gasoline and boiling range. Branched-chain compounds have relatively high octanes. The light olefins are produced in the catalytic cracker.

Current Fuel Use Patterns

The primary fuel used in the model refinery is a blend of off gases from various units within the refinery. Off gases are collected from a number of processing units, the major sources being the catalytic cracker and the catalytic reformer. The quantity and quality of the blended refinery gas varies daily, but the average quantity is 1.45 x 10^5 Nm^3/day (5.12 x 10^6 scf/day) and the average composition is

Specie	Mole, percent
H_2	29.7
C_1	32.7
C_2's	13.0
C_3's	10.4
C_4's	6.9
N_2	7.2.

The average heating value of this gas is about 37.3 MJ/Nm^3 (1000 Btu/scf). Thus, the refinery gas supplies about 5.40 x 10^6 MJ/day (5.12 x 10^9 Btu/day) of heat.

The collected off gases go to a fuel gas drum which provides for gas mixing and surge volume. Purchased natural gas is added to the fuel gas drum as needed to maintain a desired pressure level which is usually about 45 to 50 psig. The blend of refinery gas and natural gas is then distributed to the various burners in the refinery.

The model refinery manufactures asphalt from a portion of its residual oil. Due to the high demand for this product in the summer and negligible demand in winter, the refinery would vary its operations between summer and winter accordingly. The average amount of natural gas required would be 19,000 Nm^3/day (673,000 scf/day) during the summer (May through November) and 570 Nm^3/day (20,000 scf/day) during the winter (October through April). These quantities correspond to heating values of 0.21 x 10^6 MJ/day (0.02 x 10^9 Btu/day) in the winter.

During the winter, an additional 2.17 x 10^6 MJ/day (2.06 x 10^9 Btu/day) of heat would be supplied by burning residual (No. 6) fuel oil. This fuel oil has a heating value of about 43,000 kJ/kg (18,500 Btu/lb). When this fuel oil is used, it is atomized with steam and fed to the burners along with the refinery gas/natural gas blend. For the processes which are set up to burn the oil, the heat input from the oil is restricted to about 10 percent of the total heat input of the furnace. This is necessary to minimize operating problems, since the process heaters were not designed for oil.

Adding the above figures, the total heat supplied by refinery gas, natural gas, and residual fuel oil is about 6.11 x 10^6 MJ/day (5.79 x 10^9 Btu/day) in the summer and 7.60 x 10^6 MJ/day (7.20 x 10^9 Btu/day) in the winter.

Geographic Considerations

The model refinery is assumed located close to plentiful supplies of coal which could be used for the production of low-energy gas. The model refinery would also be bounded by a navigable waterway which could be used for barging coal into a gasification plant and for supplying the water needs of such a plant. The refinery could also be accessed by rail transport.

Refineries typically are located near to a number of other industrial facilities, which introduces the possibility that a single gasification plant could supply low-energy gas to this refinery plus other nearby facilities. This concept is beyond the scope of this study.

Other Considerations

The model refinery processes low-sulfur crude oil (normally less than 1 weight percent sulfur). The refinery has no sulfur plant and uses no hydrodesulfurization processes except for the removal of trace amounts of sulfur from the feed to the catalytic reformer, which is always a required operation. The products of the refinery are low in sulfur content. The residual fuel oil produced contains less than 2 weight percent sulfur.

Potential Demand for Low-Energy Gas

In considering the retrofitting of this refinery to use low-energy gas, the first priority is for replacing the purchased natural gas. The second priority is for replacing the residual fuel oil burned during the winter. Since this is low-sulfur fuel oil, it should be regarded as a premium fuel which could be used in a number of industrial facilities for which other means of controlling sulfur oxide emissions would be less practical. The residual fuel oil is also difficult to use in the existing furnaces at the refinery.

The third priority is for replacing several species in the refinery gas which have other uses for which they are better suited. One of these species is hydrogen, which can be used in hydrotreating operations in the refinery. Hydrotreating not only reduces the sulfur content of petroleum fractions, but also increases the volume of the products by adding hydrogen to them. Thus, the hydrogen can be used to produce more and cleaner liquid fuels. Hydrogen can also be marketed for other uses. The other species which could be displaced from the refinery gas are propane and butane. These are premium fuels which are normally recovered and marketed, either separately or as "liquified petroleum gas" (LPG). Propane and butane are normal products of the model refinery; the amount of these products normally recovered depends on available storage and market demand. The recovery of additional quantities of these species from the refinery gas is attractive considering the increasing price and demand for these premium fuels.

Table 21 shows the potential demand for low-energy gas at the model refinery. Based on displacing the purchased natural gas, the residual fuel oil burned, and 98 percent of the hydrogen, propane, and butane from the refinery gas, the potential demand is about 3.51×10^6 MJ/day (3.33×10^9 Btu/day) in the summer and 5.00×10^6 MJ/day (4.74×10^9 Btu/day) in the winter.

Comparison of the Model Refinery With Other Refineries

Size. The model refinery, with a crude oil capacity of about 3.97×10^6 liter/day (25,000 barrel/day), is close to the median size but less than the average size U.S. refinery. Because it is close to the median size, it is felt to be a good model with respect to size.

TABLE 21. POTENTIAL DEMAND FOR LOW-ENERGY GAS AT MODEL REFINERY

Fuel Displaced	Heating Value Demand					
	10^9 MJ/day			10^9 Btu/day		
	Summer (May-Nov)	Winter (Dec-Apr)	Annual Average	Summer (May-Nov)	Winter (Dec-Apr)	Annual Average
Purchased natural gas	0.17	0.01	0.10	0.67	0.02	0.40
Residual fuel oil burner	--	0.52	0.22	--	2.06	0.86
Hydrogen in refinery gas(a)	0.10	0.10	0.10	0.41	0.41	0.41
Propane in refinery gas(a)	0.31	0.31	0.31	1.21	1.21	1.21
Butane in refinery gas(a)	0.26	0.26	0.26	1.04	1.04	1.04
TOTAL	0.84	1.20	0.99	3.33	4.74	3.92

(a) Heating value demand based on 98 percent recovery of specie from refinery gas.

Complexity. The Nelson complexity factor for the model refinery is 5.92 and that for the U.S. refining industry as a whole is 8.88. Thus, the model refinery is less complex that the average U.S. refinery. As a result of this difference in complexity, one would expect (based on Figure 19) the unit energy consumption for the model refinery to be about 34 percent less than that of the average U.S. refinery.

Refinery Energy Consumption. A comparison of the energy consumption of the model refinery with the U.S. average value is shown in Table 22. The annual average consumption of refinery gas, natural gas, and residual fuel oil by the model refinery corresponds to about 1.55 MJ/liter crude oil (233,000 Btu/B crude oil). This is not a total energy consumption in the sense of the Bureau of Mines data(20) since it does not include coke or purchased electricity. The amount of coke consumed as fuel is difficult to estimate because this includes the coke deposited on catalysts in process units and then burned off, such as is done in catalytic crackers. The amount of electricity used for process units at the model refinery must be included to determine total energy use. In order to obtain an approximate comparison with the Bureau of Mines data(20), one can add to the known energy consumption of the model refinery the average values for coke and electricity for the state (or group) in which the model refinery is located. This gives a total energy consumption of about 2.20 MJ/liter crude oil (330,000 Btu/B crude oil).

For comparison with the model refinery, the U.S. average energy consumption has to be adjusted for the differences in time (fuel cost) and complexity. Using Nelson's correlation (Figure 19) to correct the U.S. average value to the time and complexity of the model refinery cases gives a total energy consumption of about 2.38 MJ/liter crude oil (358,000 Btu/B crude oil). This agrees reasonably well with the value of 2.20 MJ/liter cited above. Thus, the total energy consumption of the model refinery appears to fall in line reasonably well with other industry data when the effects of the pertinent variables are properly considered.

TABLE 22. COMPARISON OF ENERGY CONSUMPTIONS FOR MODEL REFINERY WITH U.S. AVERAGE VALUES

	Net Energy Consumption, J/litre (10^3 Btu/B crude)[a]			
	1973 U.S.	Model Refinery, 1975		
Energy Source	Refinery Average	May-Nov	Dec-Apr	Annual Average
Crude oil	0.331 (0.05)			--
Distillate fuel oil	47.1 (7.09)			--
Residual fuel oil	404.7 (60.98	0.0 (0.0)	548.2 (82.6)	228.3 (34.4)
Liquefied petroleum gas	59.5 (8.96)			--
Natural gas	1619.2 (243.99)	184.5 (27.8)	5.3 (0.8)	109.5 (16.5)
Refinery gas	1568.7 (236.38	1206.5 (181.8)	1206.5 (181.8)	1206.5 (181.8)
Petroleum coke	585.1 (88.17)			--
Coal	11.6 (1.74)			--
Purchased electricity	116.7 (17.58)			--
Purchased steam	59.6 (8.98)			--
TOTAL	4472.5 (673.92)[b]			1544.3 (232.7)[c]
Refinery Complexity	9.24		5.92	

(a) MJ/liter crude = (1.591) (10^3 Btu/B crude).

(b) Adjusting from 1973 to 1975 and from complexity 9.24 to 5.92 using Figure 3 yields

$$(6.73.92)(\frac{353}{664}) = 358 \times 10^3 \text{ Btu/B.}$$

(c) Does not include coke or purchased electricity. Adding average values of these for state (or group) of model refinery gives total of 330 10^3 Btu/B.

Types of Fuels Used. The model refinery is heavily dependent upon the refinery gas as an energy source. While this is a major energy source in most refineries, there are many refineries in which it is not nearly so dominant as was assumed for the model refinery. The data in Table 18 indicate that, on a national basis, refinery gas provides about 35 percent of the total energy needs of refineries. This compares with about 55 percent for the model refinery (including the estimated coke and electricity).

For the model refinery, low-sulfur crude oil is processed, no other hydrotreating operations are used, and, hence, much of the hydrogen from the catalytic reformer can be used for fuel. In many other refineries, the crude oil will contain more sulfur, more of the hydrogen will be required for hydrotreating operations, and, hence, less of the fuel needs will be satisfied with refinery off gases. Polymerization is used in the model refinery, but alkylation is much more widely used in other refineries. Therefore, most of the C_2-C_4 compounds in the refinery gas of the model refinery are assumed to be saturated hydrocarbons (paraffins).

Land Area. The model refinery occupies a total of about 32 acres, of which only about 3 acres are used for the processing equipment. As can be seen from Figure 23, the area per unit throughput for the model refinery is somewhat less than the general correlation would indicate. Thus, the model refinery is probably somewhat more compact than many other refineries.

Access to Waterway. The model refinery is located on a navigable waterway, and this is true for most other refineries as well. Refineries, crude oil is received and refined products are shipped by tankers and/or barges.

Gasification Plant Design

Due to the low overall energy demand of the model refinery, an air-blown fixed-bed, Wellman-Galusha gasification system was selected for study. The gasification plant would supply about 5.00 x 10^6 MJ/day (4.74 x 10^9 Btu/day) in the winter and 3.51 x 10^6 MJ/day (3.33 x 10^9 Btu/day) in the summer. Figure 24 shows the flow sheet for the Wellman-Galusha gasification plant. The mixture of refinery waste gas at 39.6 MJ/Nm^3 (1062 Btu/scf) and low-energy gas from the Wellman-Galusha at 6.26 MJ/Nm^3 (168 Btu/scf) would have a heating value of about 9.84 MJ/Nm^3 (264 Btu/scf) in the winter and 8.72 MJ/Nm^3 (234 Btu/scf) in the summer. A complete material balance for this plant is given in Appendix B. Table 23 summarizes the pertinent characteristics of the refinery model gasification plant.

TABLE 23. GASIFICATION PLANT DESIGN FOR REFINERY MODEL

Gasifier - Wellman-Galusha (3 units)

Desulfurization - Stretford

Maximum gas production rate - 5.0 x 10^6 MJ/day
(4.77 x 10^9 Btu/day)

Gas high heat value - 6.619 MJ/Nm^3 (168 Btu/scf)

Coal consumption - 228 metric ton/day
(252 ton/day)

Efficiency - 76.7 percent

The coal selected for use in this system was an Eastern bituminous-type coal with 6 percent moisture, 8 percent ash, and sulfur content of 3.9 percent. The free-swelling index of this coal is about 5, dictating the use of an agitator-type fixed-bed gasifier. A complete analysis of the coal is given in Table 24.

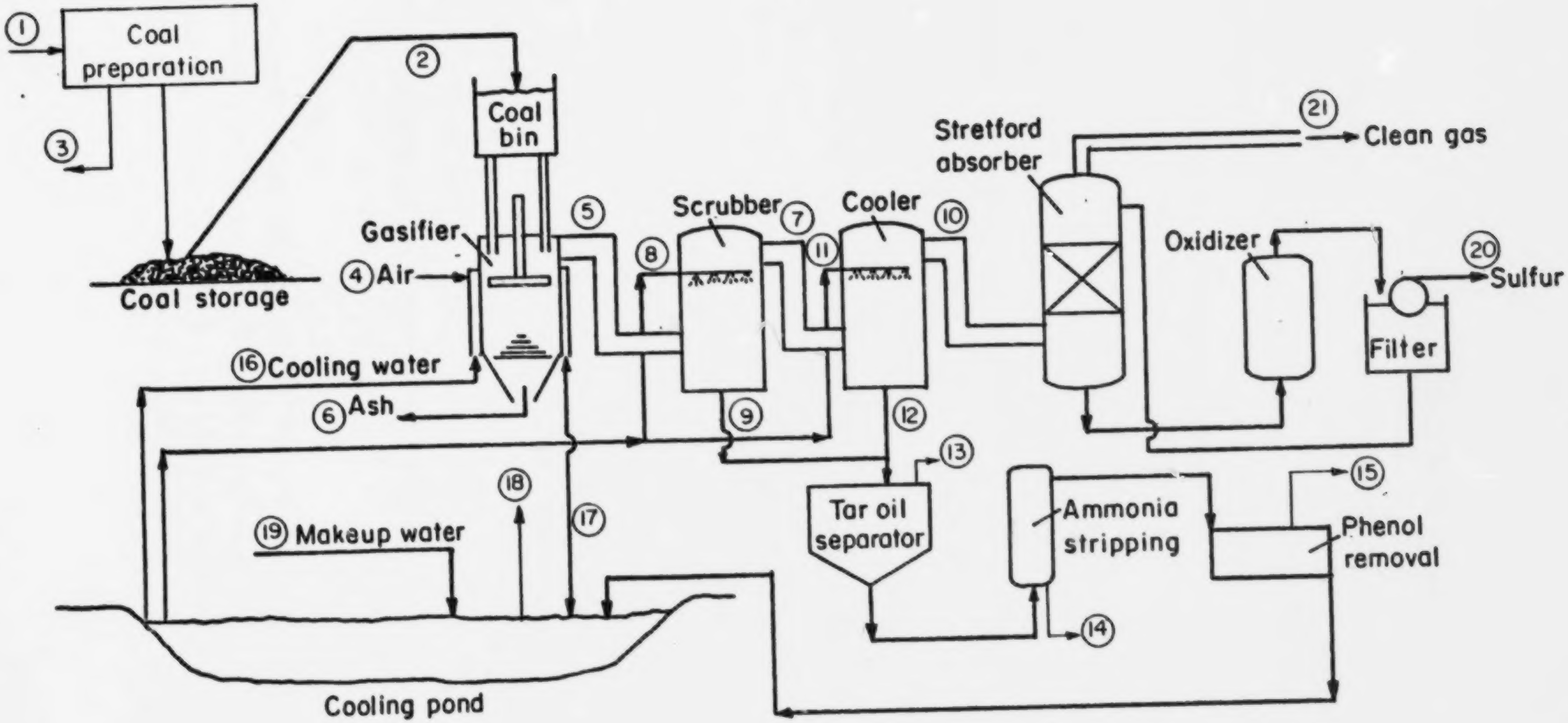

①
Coal preparation
②
③
Coal storage
Coal bin
Gasifier
④ Air
⑤
⑯ Cooling water
⑥ Ash
⑧
Scrubber
⑦
⑨
⑪
Cooler
⑩
⑫
⑬
Tar oil separator
Stretford absorber
㉑ Clean gas
Oxidizer
⑳ Sulfur
Filter
Ammonia stripping
⑭
⑮
Phenol removal
⑱
⑰
⑲ Makeup water
Cooling pond

FIGURE 24. FLOW SHEET FOR THE WELLMAN-GALUSHA GASIFICATION PLANT
(See Appendix B for complete material balance.)

TABLE 24. REFINERY MODEL PLANT COAL ANALYSIS

Proximate	Wt. Percent
Moisture	6.1
Volatile Matter	32.7
Fixed Carbon	48.3
Ash	8.4
Ultimate Analysis	
Hydrogen	4.8
Carbon	68.0
Nitrogen	2.1
Oxygen	6.8
Sulfur	3.9
Ash	8.4
HHV (Btu/lb)	13690
Grindability	60
Free-Swelling Index	5

Due to the small size of the gasification plant it was not felt practical to install coal preparation facilities, therefore, crushed-sized coal would be purchased from the mine and stored at the gasification plant. The gas plant would consist of three 10-ft diameter Wellman-Galusha units capable of producing a total of 5.03 x 10^6 MJ/day (4.77 x 10^9 Btu/day) of fuel gas with a heating value of 6.26 MJ/Nm^3 (168 Btu/scf). The coal consumed would be about 278 metric ton/day (252 ton/day) and the overall thermal efficiency of the plant would be 76.7 percent. The raw gas from the gasifier is processed directly through a scrubber for the removal of tars, oil, phenols, and ammonia, and then through a cooler section where additional ammonia, tars, and other condensible constituents are removed. The gas is then fed into a Stretford-type desulfurization system which oxidizes sulfur compounds to elemental sulfur in solution, eliminating the need for a Claus plant. The final gas product would contain 300 ppm of sulfur, or less, and would be mixed with refinery gas and distributed to the various processes in the refinery.

Figure 25 shows an overall plot plan of the refinery with gasification and coal storage facilities and also the required cooling pond. The processing facilities of the refinery itself occupy about 16,200 m^2 (4 acres) of ground, and storage capacity requires an additional 93,000 m^2 (23 acres) of ground. The gasification plant for the refinery is estimated to require about 4,050 m^2 (1 acre) of ground with an additional 4,050 m^2 (1 acre) required for a cooling pond. Coal-storage facilities for 1-month supply of coal would require an additional 4,050 m^2 (1 acre).

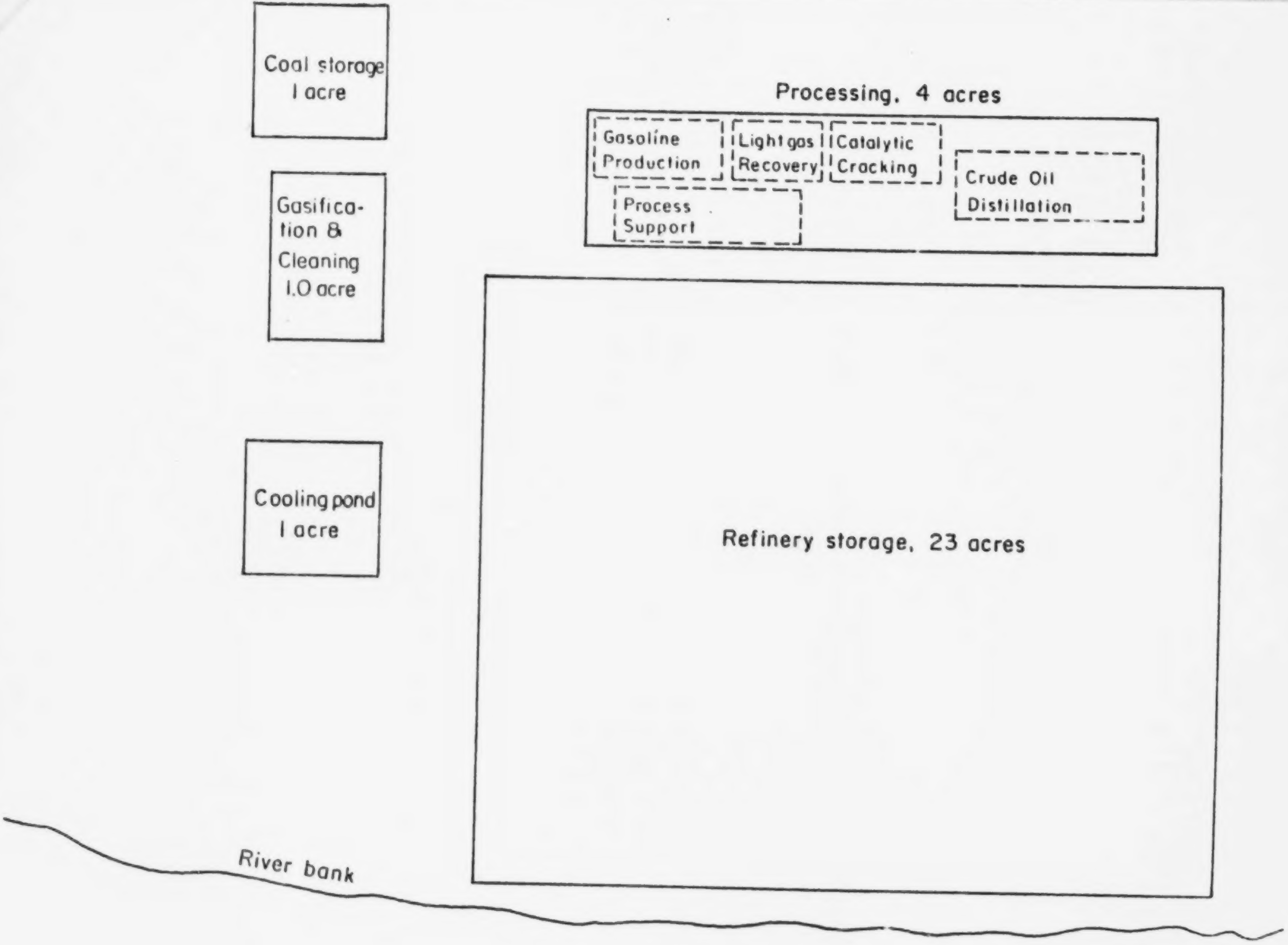
Potential Impact of Low-Energy Gas

The petroleum refining industry is a promising candidate for retrofitting to the use of low-energy gas because the consumption of energy, and particularly natural gas, by the industry is high and because much of the industry is located in regions of high coal availability. The industry includes a wide range of refinery sizes and energy requirements. The model refinery used in this study is somewhat small when one considers the economic justification of a coal-gasification facility to serve a single refinery. It is important to look at some of the larger refineries in the United States in order to appreciate the impact which the use of low-energy gas from coal could have on the nation's refining industry.

Table 25 lists the 24 largest petroleum refineries in the United States. For each of these refineries, the table gives the Nelson complexity factor and estimates of the consumptions of energy in the form of natural gas and oil. The latter were obtained by

(1) Determining the unit energy consumption (all sources) from the complexity using Nelson's correlation (Figure 19)

(2) Multiplying the above by the fraction of the total energy supplied by natural gas and oil, using the Bureau of Mines data for the state or state group in which the refinery is located (Table 18).



Coal storage
1 acre
Gasification & Cleaning
1.0 acre
Cooling pond
1 acre
Processing. 4 acres
Gasoline Production
Light gas Recovery
Catalytic Cracking
Crude Oil Distillation
Process Support
Refinery storage, 23 acres
River bank

FIGURE 25. REFINERY PLOT PLAN

TABLE 25. CAPACITIES AND ESTIMATED ENERGY CONSUMPTION(17) OF LARGEST REFINERIES IN THE UNITED STATES

State	Company	City	Coal Availability	Crude Oil Capacity,(17) 10^3 B/CD	Nelson Complexity	Unit Natural Gas and Oil Consumption		Total Natural Gas and Oil Consumption	
						MJ/liter crude	10^3 Btu/B crude	10^9 MJ/day	10^9 Btu/day
(1) Louisiana	Exxon	Baton Rouge		445	9.45	396	249	28.1	111
(2) Texas	Texaco	Port Arthur	High	406	8.69	500	314	32.1	127
(3) Texas	Exxon	Baytown	High	400	12.99	749	471	47.6	188
(4) Indiana	Amoco	Whiting	High	360	9.26	366	230	21.0	83
(5) Texas	Amoco	Texas City	High	333	10.23	590	371	31.4	124
(6) Texas	Mobil	Beaumont	High	325	9.18	528	332	27.3	108
(7) Texas	Gulf	Port Arthur	High	312	10.24	590	371	29.3	116
(8) Texas	Shell	Deer Park	High	294	9.77	563	354	26.3	104
(9) Illinois	Shell	Wood River	High	283	10.15	318	200	14.4	57
(10) Louisiana	Cities	Lake Charles		268	9.52	399	251	17.0	67
(11) New Jersey	Exxon	Linden		265	8.15	325	204	13.7	54
(12) Louisiana	Shell	Norco		240	7.27	304	191	11.6	46
(13) Mississippi	Std (Calif)	Pascagoula		240	6.99	333	209	12.7	50
(14) California	Std (Calif)	El Segundo		230	7.94	320	201	11.6	46
(15) Texas	Atlantic-Richfield	Houston	High	213	10.12	584	367	19.7	78
(16) Illinois	Marathon	Robinson	High	195	5.84	183	115	5.6	22
(17) California	Std (Calif)	Richmond		190	12.45	504	317	15.2	60
(18) California	Atlantic-Richfield	Carson		185	8.28	334	210	9.9	39
(19) Pennsylvania	Atlantic-Richfield	Philadelphia	High	185	5.75	231	145	6.8	27
(20) Louisiana	Gulf	Belle Chasse		180.4	8.17	342	215	9.9	27
(21) Illinois	Mobil	Joliet	High	175	8.62	270	170	7.6	30
(22) Pennsylvania	Gulf	Philadelphia	High	174.3	7.83	315	198	8.9	35
(23) Ohio	Sohio	Lima	High	168	7.50	229	144	6.1	24
(24) Pennsylvania	Sun	Marcus Hook	High	165	13.73	554	348	14.4	57

Table 25 also indicates which refineries are located in states considered as having a high coal availability (production or reserves).

Among the refineries listed in Table 25 are

- Seven refineries having estimated natural gas and oil consumption greater than 83 x 10^6 MJ/day (79 x 10^9 Btu/day) and located in states having high coal availability.
- Nine refineries having estimated natural gas and oil consumption of 21-83 x 10^6 MJ/day (20-79 x 10^9 Btu/day) and located in states having high coal availability
- Eight refineries located in states not having high coal availability.

With regard to coal availability, it should be noted that there may be cases in which transporation of coal from a nearby state is feasible. On the other hand, there may be cases in which coal reserves in a given state are not feasible for use at a site within the same state but are fairly far away.

For these large refineries, the estimated natural gas and oil consumptions are high enough to justify on-site coal gasification facilities. It appears that there are quite a number of refineries in the United States for which the energy needs and locations are such that retrofitting them to use low-energy gas from coal could make considerable sense. The impact of this option upon the petroleum refining industry could be quite significant.

Burners and Furnaces in a Refinery Plant

This section describes typical furnaces and burners used in a refinery plant similar to that described in this study and the possibilities of converting these processes to low-energy gas.

Burners

In the case of the refinery discussed in this study, because of the relatively small fuel needs, the use of a Wellman-Galusha air-steam gas producer is proposed. This gas would be mixed with refinery gas from which marketable components such as hydrocarbon, propane, and butane had been stripped leaving CH_4, C_2H_6, and N_2 as components in a gas of 41.8 MJ/Nm^3 (1062 Btu/scf) high heating value (HHV). The HHV of the fuel mixture is 10.4 MJ/Nm^3 (264 Btu/scf) during the summer (when the mixtures contain 10.8 percent by volume refinery gas). The HHV during the winter is 9.25 MJ/Nm^3 (235 Btu/scf) (when the fuel gas contains 7.5 percent by volume of refinery gas). The summer and winter Wobbe numbers would be 291 and 258. The flash-back velocity gradient at stoichiometric and the heat release rate at stoichiometric are very slightly below the values given in Table A-2 for Wellman-Galusha gas, and somewhat above those for natural gas.

For the fuel mixture the flash-back velocity gradient times the higher heating value, a probably important criterion for nozzle-mix type burners, varies from 591 to 512 MJ/Nm^3-sec (15×10^4 to 13×10^4 Btu/scf-sec). These values are far below the value for natural gas, but still an improvement over Wellman-Galusha gas.

For the typical refinery considered in this study, Table 26 lists characteristics of the furnaces used for process heating and steam raising. Figure 26 shows an inspirating burner used on refinery furnaces. Figure 27 shows the burner used in refinery boilers.

On changeover to the mixed fuel from natural gas, it is probable that all the furnace burners would have to be changed to gas burners of the general type shown in Figure 28, when sufficient draft is available. When sufficient draft was not available, exhaust fans could be added, or nozzle-mix burners with blowers would be used.

Because of the low-heating values of the gas, the burners in boilers would require changing. (See discussion of secondary steel plant boilers.) One boiler manufacturer would recommend a vortex burner for the low heating value gases. They would also recommend replacing the multiple

TABLE 26. FURNACES IN A SMALL REFINERY

No.	Type	Design Capacity, 10^3 MJ/hr	Design Capacity, (10^6 Btu/hr)	Temperature, F (F) Stack	Temperature, F (F) Furnace	O_2 Percent	Efficiency, Percent	Fuels
1	Preflash reboiler	8.2	(7.8)	516 (960)	668 (1235)	4.5	69.6	Gas, No. 6 oil
2	Crude heater	20.3	(19.5)	504 (940)	757 (1395)	2.0	71.5	Gas, No. 2 oil
3	Vacuum tower heater	14.2	(13.5)	574 (1065)	689 (1270)	4.5	66.8	Gas, oil
4	Light oil heater	29.7	(28.2)	493 (920)	654 (1210)	3.5	71.6	Gas, No. 6 oil
5	Tar stripper heater	38.8	(36.8)	643 (1190)	663 (1225)	4.8	63.0	Gas, No. 6 oil
6	Unifier heater	10.5	(10.0)	609 (1128)	-- --	2.6	64.0	Gas
7	Platforming heater	21.1	(20.0)	-- --	649 (1200)	4.8	--	Gas
8	Raw oil heater	5.9	(5.6)	527 (980)	677 (1250)	9.2	54.0	Gas
9	Boilers	47-65	(45-62)	-- --	-- --	-	--	Gas, No. 6 oil

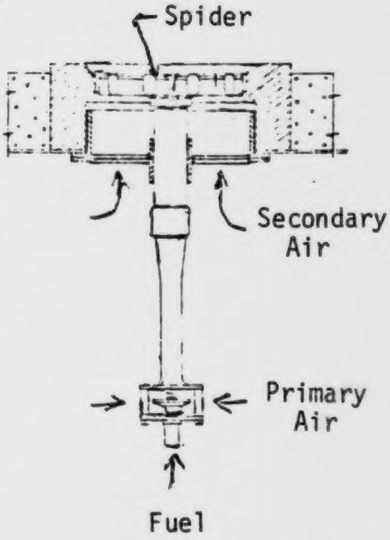

Spider
Secondary Air
Primary Air
Fuel

FIGURE 26. ZINK VPM VERTICAL GAS BURNER FOR HIGH HYDROGEN GAS

Spider with radial arms distributes primary air-fuel mixture evenly over secondary air stream. No adjustment needed in shifting from start-up gas to high hydrogen fuel.

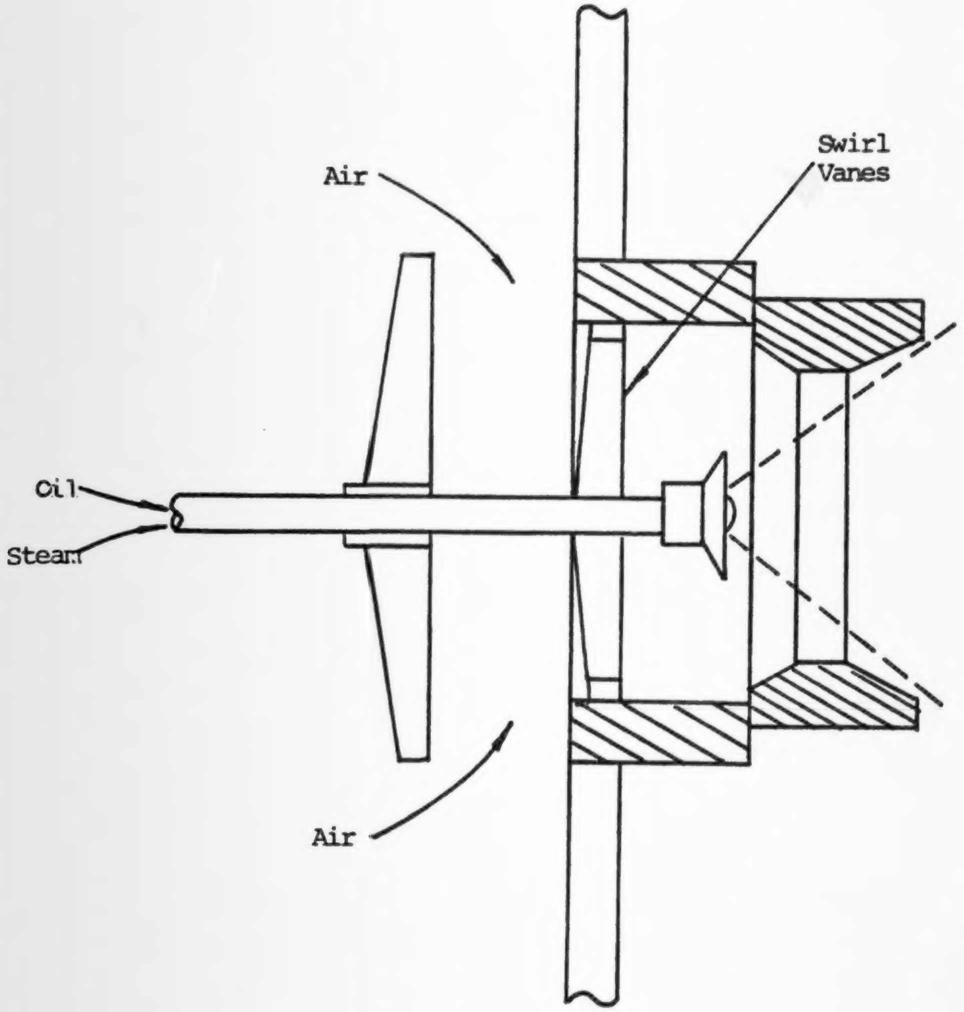

Air
Swirl
Vanes
Oil
Steam
Air

FIGURE 27. REFINERY BOILER BURNER

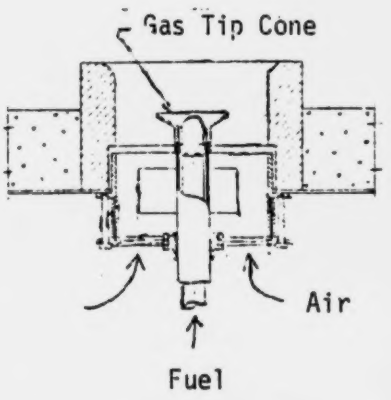


FIGURE 28. ZINK VYR VERTICAL GAS BURNER FOR PROCESS HEATERS

Burner designed to use raw gas at appreciable pressure, and natural draft to supply air. Has a high turndown ratio and can use a wide variety of gases. Gas-tip cone is perforated with slots to permit passage of air into recirculation zone.

burners with a single large capacity burner. This would cut down cost of replacing air ducting. However, because of the high cost of field work, it is quite possible that the replacement of the entire boiler-burner systems with new package units would be the most economical approach.

To summarize, it is probable that all the burners in a refinery might have to be replaced when a change is made from the natural gas. Further, it may be most economical to replace the boilers with new package boilers rather than attempt to make field changes on their burners.

V. CONSIDERATIONS IN DISTRIBUTING LOW- AND INTERMEDIATE-ENERGY GAS IN INDUSTRY

Volume and Pressure Considerations

Industrial gas distribution systems are often intricate and extensive. The model steel plant in this study would have approximately 9144 m (30,000 ft) of gas piping with diameters ranging from 38 to 254 mm (1-1/2 to 10 inches). The refinery model would have approximately 762 m (2500 ft) with sizes ranging from 25 to 152 mm (1 to 6 inches) in diameter. These piping systems would be carbon steel with some brass valves and fittings. Natural gas distribution systems are commonly rated at about 1030 kPa gage (150 psig). In most plants in the two industries considered in this study, however, natural gas would be distributed at much lower pressures of about 276 to 345 kPa gage (40 to 50 psig).

A schematic of an industrial piping system is shown in Figure 29. The gas is supplied to the system at some supply pressure, Ps, and exits the system at the burner at pressure, P_E. The difference between Ps and P_E is the pressure drop through the system which for turbulent flow is proportional to the gas density (ρ) times the square of the velocity (V). Prior to being admitted to the burner, the exit pressure, P_E, is further reduced by an orifice to a pressure normally less than 6.9 kPa gage (1 psig).

Because natural gas is often distributed at much less than the design pressure of the distribution system, it is useful to look at the possibility of using the same system for a lower energy gas. The governing equation relating the supply and exit pressures for two gases (1 and 2) assuming the same energy supply rate for both cases is

$$P_{s2}^2 - P_{E2}^2 = \frac{W_1^2}{W_2^2} (P_{s1}^2 - P_{E1}^2)$$

where W = Wobbe number = HHV $\sqrt{\rho}$ at standard conditions. An extreme, simplified case would be where $P_{E1} = P_{E2} = 0$. The equation then reduces to

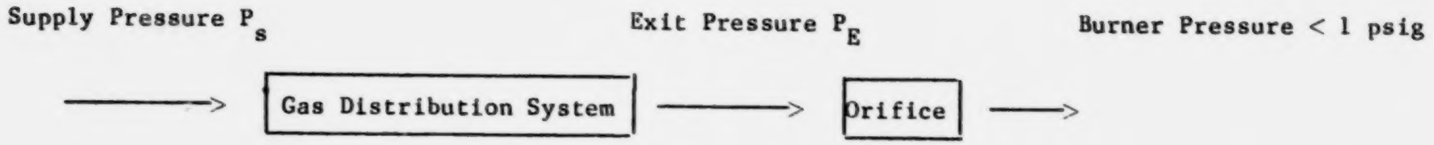

Supply Pressure P_s
Gas Distribution System
Exit Pressure P_E
Orifice
Burner Pressure < 1 psig

FIGURE 29. INDUSTRIAL GAS DISTRIBUTION SYSTEM

$$P_{s2} = \frac{W_1}{W_2} P_{s1} \quad .$$

Figure 30 shows this relationship for 3 cases: Wellman-Galusha gas, Wellman-Galusha gas mixed with refinery gas for the refinery model, and Koppers-Totzek gas for the steel plant model.

As can be seen in Figure 29, for a natural gas supply pressure of 207 kPa gage (30 psig), a pressure of 1070 kPa gage (155 psig) would be necessary for the Koppers-Totzek gas in the steel mill model and over 1380 kPa gage (200 psig) would be necessary for the Wellman-Galusha gas and refinery gas. Both the steel mill model plant and the refinery model plant were assumed to have a natural gas supply pressure of from 276 to 345 kPa gage (40 to 50 psig). It would be concluded, therefore, that using the existing distribution system would require pressures that would exceed the design pressure of the existing system. It would be assumed that at least part or all of the gas distribution system would have to be replaced. The required pipe size would depend on the pressure at which the gas is supplied. If the gas were supplied at the same pressure as the natural gas and the total pressure drop through the system were kept constant, then the required pipe areas for two gases are related by

$$\frac{A_1}{A_2} = \frac{W_1}{W_2} \quad .$$

For the three cases shown in Figure 29, the area ratios would be as shown in Table 27.

TABLE 27. REQUIRED PIPE SIZE FOR GAS DISTRIBUTION*

	HHV Btu/scf	W	Pipe Area Ratio	Pipe Diameter Ratio
Natural Gas	973	1244	1	1
Wellman-Galusha Gas	168	183	6.8	2.61
Wellman-Galusha Refinery Gas Mixture	235	256	4.8	2.19
Koppers-Totzek Steel Mill Gas	286	338	3.7	1.92

* Assuming the same supply pressure and heat delivery rate.

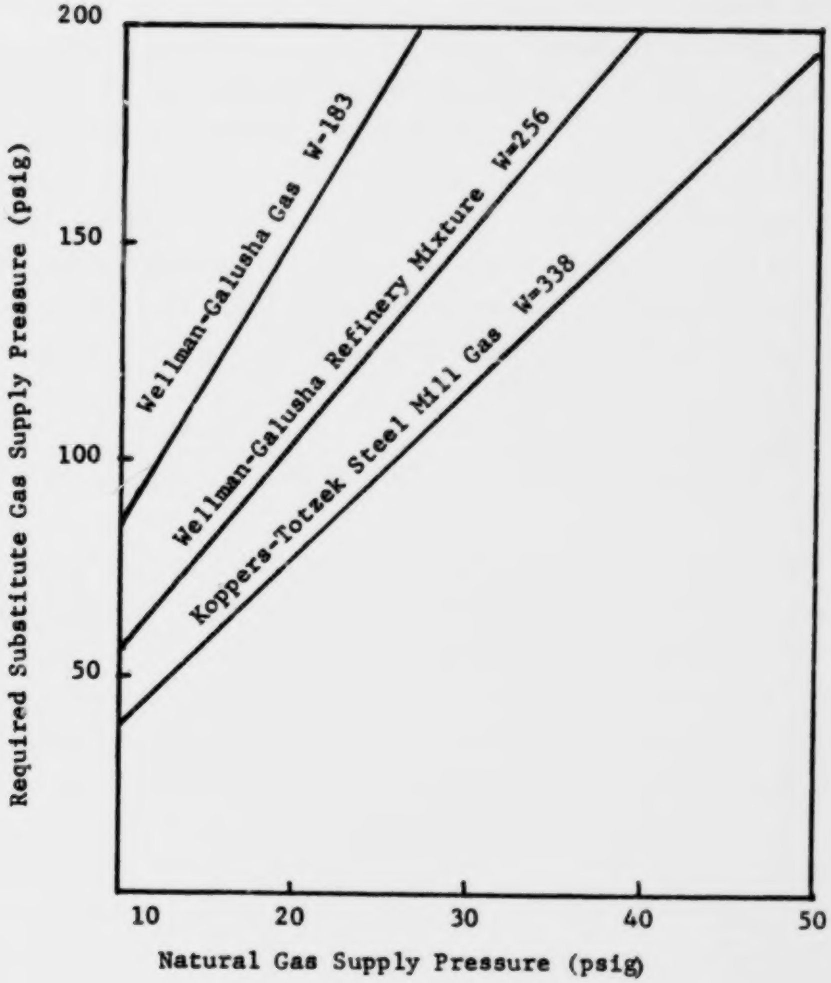


FIGURE 30. REQUIRED GAS SUPPLY PRESSURE FOR SUBSTITUTING GAS FROM COAL FOR NATURAL GAS IN AN EXISTING DISTRIBUTION SYSTEM

The size of pipe and its cost would have to be weighed against the available space and costs of compression. Compression could require a significant amount of energy depending on the final gas pressure. The most efficient way to compress the gas is with intercooling in an isothermal process. Most large compression systems use intercooling. The other extreme is adiabatic compression where no heat is transferred from the gas as it is compressed. Figures 31 and 32 show the power requirements for both adiabatic and isothermal compression for the steel and refinery plant models, respectively. Compression of fuel gas to 6.9 x 10^5 Pa (100 psig) ($P_2/P_1 = 7.8$) in both model industry cases would require 0.40 x 10^6 MJ/day (0.38 x 10^9 Btu/day) for the steel mill model (which is 1.8 percent of the total energy in the clean gas) and 0.15 x 10^6 MJ/day (0.14 x 10^9 Btu/day) for the refinery model (which is 2.9 percent of the total energy in the clean gas).

Corrosion Considerations on Substituting Low- or Intermediate-Energy Gas for Natural Gas

Potential corrosion problems in gas distribution systems and process equipment resulting from the substitution of low- or intermediate-energy gas from coal for natural gas are also an important consideration. Corrosive constituents in the produced fuel gas can increase degradation of carbon steel, brass, and other materials found throughout fuel systems. Specific interest is given here to retrofitting a steel plant to fuel gas; however, the discussion has general applicability to a variety of industrial processes.

While the composition of gases produced by coal gasifiers is somewhat unique, a broad experience exists for handling of corrosive gases from other sources, e.g., coke oven gas, sour gases from petroleum production, gases generated in chemical processes, and refinery industries. Experience with distribution of town gas, used extensively in Europe, is directly applicable. The approach taken in this study was to identify the corrosive species in fuel gas and, where possible, to establish acceptable limits for distribution. Also, corrosion mitigation and monitoring procedures were reviewed.

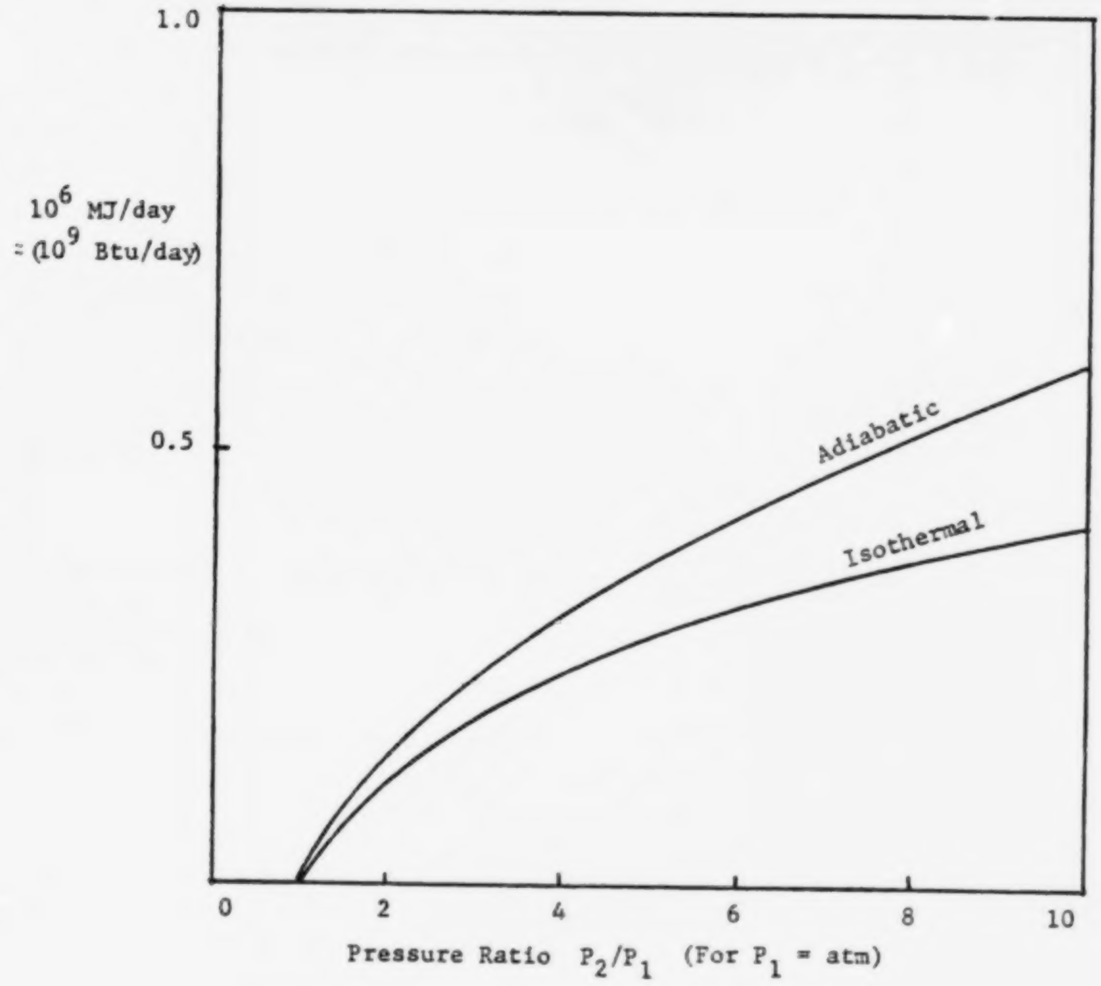


FIGURE 31. COMPRESSION POWER FOR STEEL MILL MODEL GAS SUPPLY

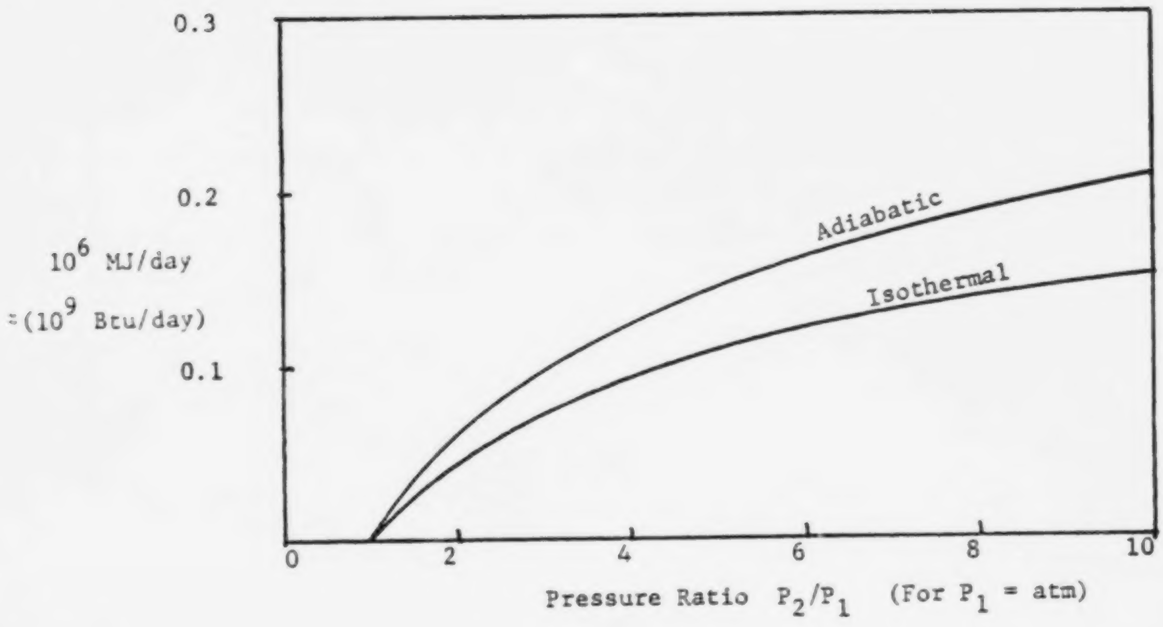


FIGURE 32. COMPRESSION POWER FOR REFINERY MODEL GAS SUPPLY

Corrosive Species in Low- and Intermediate-Energy Gas from Coal

A variety of materials is used in the distribution and usage of fuel gas. Carbon steel for pipes and fittings is the most prevalent material with lesser amounts of brass found in valves and high-alloy steels and nickel alloys in process equipment. Constituents in fuel gas in the presence of water support corrosion of these materials. Of primary concern are conditions resulting in general corrosion, but those which promote stress-corrosion cracking are also considered.

Constituents of fuel gas can be divided into three groups: corrosive, inhibitive, and inert. Carbon dioxide (CO_2), hydrogen sulfide (H_2S) and other sulfur-containing species, ammonia (NH_3), and hydrogen cyanide (HCN) promote corrosion. Carbon monoxide (CO) inhibits corrosion; whereas hydrogen (H_2), methane (CH_4), and nitrogen (N_2) do not significantly affect corrosion. The acid gases, carbon dioxide and hydrogen sulfide, readily corrode mild steel. Copper alloys are corroded by sulfur compounds and are susceptible to corrosion or stress-corrosion cracking in the presence of ammonia. Nickel and nickel alloys are corroded by sulfur compounds. Although the effect of a given species on corrosion is generally known, the corrosivity of a mixture of gases is not readily predictable because of complex interaction and temperature effects.

Of all the constituents in fuel gas, hydrogen sulfide is the most deleterious because even small amounts can greatly accelerate corrosion. An early study of the corrosion of steels by natural gas containing traces of H_2S recommends that H_2S content be controlled to less than 2.28 mg/m^3 (0.1 grain per 100 cubic feet of gas)(26). Corrosion is not severe in the absence of water. Corrosion of steel in refinery condensing systems was found to increase with sulfide concentration(27). Inhibitor treatment and pH control(28) were necessary to control corrosion of steel in storage of high-pressure sour gas (13 percent H_2S, 5 percent CO_2). Monel and Inconel alloys were substituted for austenitic stainless steel in special equipment operating at ambient temperatures. Many other instances are recorded in which severe corrosion problems arose from handling of moist hydrogen sulfide containing gases.

In addition to general corrosion, stress-corrosion cracking (SCC) is promoted by H_2S. Susceptibility to SCC for a range of ferrous materials increases as hardness increases. In sour gas service, most failures of tubular components occurred with alloys, the hardness of which exceeded R_C22(29). Field failure data and laboratory studies form the basis for NACE's publication IF 166, "Sulfide Cracking Resistant Metallic Materials for Valves for Production and Pipeline Service" which recommends R_C22 as the maximum hardness level for this service. The recommendation has seen much broader application than just for valves.

Based upon laboratory data, 0.001 atmosphere was chosen as the critical partial pressure of H_2S at which SCC will occur(30,31). Under more severe conditions, higher temperature and pressure, the value is lower still(32). The point to be made is that even small amounts of H_2S can promote SCC.

Carbon dioxide dissolves in water to form carbonic acid, a corrosive agent to mild steel. Corrosion rates in excess of 100 mils per year have been observed for partial pressures of approximately 690 kPa (100 psia). Obrect(33) identified CO_2 as a major corrodent in steam-condensate systems. It is also recognized as a primary contributor to corrosion in handling of sour gases. A rule-of-thumb for natural gas transmission is that no special corrosion mitigation procedures are required for partial pressures of CO_2 below 35 kPa (5 psia). This level is not absolute as evidenced by a steady lowering of the acceptable limit over the years. Presence of both H_2S and CO_2 lowers the tolerable limits of each g[illegible]

Mixtures of carbon dioxide-carbon monoxide-water were shown to promote SCC of a high-strength steel. Steel specimens failed in 65 percent CO - 35 percent CO_2 mixtures at total pressure as low as 2 atmospheres at 20 C(34). This system has also resulted in SCC of mild steed in town gas composition(35). Recent work at Battelle has shown SCC of mild steel to occur in CO_2-CO-CH_4-H_2O at CO_2 and CO partial pressures of 6.9 kPa (1 psia) and less. All three constituents (CO_2, CO, and H_2O) must be present to support SCC.

Ammonia accelerates the corrosion of mild steel, but is presence in fuel gas is of greater consequence because of its deleterious effect on copper and copper alloys. Stress-corrosion cracking of many of the copper alloys is readily promoted by NH_3, even at trace levels. At high concentrations, general corrosion of copper alloys is a serious problem.

Copper alloys are also corroded by sulfur-bearing compounds. Nickel and nickel alloys are susceptible to sulfidation in aqueous phase and at high temperatures. The latter is of concern when burning sulfur-containing fuel.

Other constituents of fuel gas can participate in corrosion processes, but the primary contributions to corrosivity of fuel gas are made by species discussed above: hydrogen sulfide, carbon dioxide, and ammonia.

In addition to corrosive gases, fuel gas contains condensable tars and ash, which can cause plugging and blockage if not controlled. A beneficial effect of condensable organics is that they can coat the metal surfaces and retard corrosion.

Mitigation and Monitoring of Corrosion by Fuel Gas

In the above section, it was shown that raw fuel gas contains several species which promote corrosion of materials commonly found in gas distribution systems and processes equipment burning fuel gas. Here, procedures to mitigate and monitor corrosion by fuel gas are discussed. "Control of Internal Corrosion in Steel Pipelines and Piping Systems", NACE Standard RP-01-75, presents recommended practice for corrosion control of pipeline systems, including gas transmission and gas distribution systems. Relevant portions of the recommended practice are presented below with experience from comparable service conditions, namely, transport of coke oven and town gas, transmission of natural gas, and handling of sour gases during production.

Corrosion control can be achieved in this service by several procedures: (1) elimination of corrosive species in the fuel gas, (2) use of corrosion inhibitors, (3) application of coatings, and (4) substitution of more corrosion-resistant material. Choice of procedure is made on the basis of economics and ease of application to a specific problem area.

In view of the number of corrosive species present in fuel gas and the variety of materials in contact with the gas, removal of water provides the most general means to control corrosion. In the absence of water, corrosion throughout the distribution system would be negligible. (Corrosion in town gas systems in Europe was controlled by removal of water and desulfurization(36).) Water can be removed by water separators, by refrigeration, or by dehydrators. Various types of dehydrators are available including glycol and desiccant. Using these means, the dewpoint of the gas is maintained below service temperatures to prevent condensation in the system. Commercial units are available to dehydrate large volumes of gas.

It may be advantageous to remove other corrosive constituents in addition to water. Commercial processes are available to remove acid gases, ammonia, and other corrosive species. Removal of sulfur prior to use of fuel gas decreases corrosion throughout the system (in addition to eliminating the need for flue-gas clean-up units).

Addition of corrosion inhibitor can be used in conjunction with other corrosion control procedures. Several types of inhibitors are available for either continuous or batch application. Filming inhibitors are effective for gas distribution systems. Application of protective coatings is not seen to be necessary for the bulk of the piping system, but it can be beneficial in specific areas.

In process equipment when a specific corrosion problem is identified, selection of a more corrosion resistant material may provide a ready solution. For example, nickel and high-nickel alloys are susceptible to sulfidation and are not recommended for use with high temperature sulfur bearing gases. Alloys resistant to sulfidation should be used.

The need for corrosion mitigation and the evaluation of its effectiveness are determined by analysis of corrosion monitoring data. The level of sophistication required is determined in part by the consequences of a failure. A leak in a fuel gas system is less tolerable than a leak in a natural gas system, because, in the addition to fire and explosion hazards, noxious carbon monoxide would escape. Prior to conversion to fuel gas, the entire system should be inspected and a list of materials throughout the system compiled. Any questionable components, because of present condition or known corrosion susceptibility, should be replaced. A sample of each type of component in the system should be reinspected periodically for corrosion damage after conversion to fuel gas. These inspections can be supplemented by data from corrosion coupons and probes installed throughout the system as necessary. Analysis of gas, residue, and deposits found in the system also provides valuable information. Experience gained following conversion to fuel gas will dictate the frequency and amount of inspection required.

Handling of fuel gas presents similar corrosion problems to those of handling coke oven gas, i.e., a variety of acid gases and other corrosive components are produced in a moist gas. Corrosion control practiced varies with the severity of corrosion problems experienced at different plants. Except for special instances, distribution systems of carbon steel have provided good service. For mitigation, where corrosion was excessive, the coke oven gas was either dried or partially dried and desulfurized. Low corrosion rates of carbon steel have been observed in some moist coke oven gas service with no applied conversion control. These low rates were attributed to condensable hydrocarbons coating the steel surface. Austenitic stainless steel has been used successfully to carry moist coke oven gas. However, it must be recognized that austenitic stainless steel, particularly in the sensitized condition, is susceptible to SCC in presence of polythionic acid(37), chloride, or fluoride(38). Polythionic acid and chloride can form from, or are found in, the environment, while fluoride can result from use of some welding fluxes.

Internal corrosion of natural gas transmission lines is controlled primarily by dehydration of the gas and inhibitor treatment. Inhibitor can be injected continually or by batch treatment in a pigging operation. Monitoring of internal corrosion in pipelines transporting natural gas containing CO_2 and H_2S was recently reviewed(39). Corrosion data obtained on an operating system are presented for corrosion coupons, hydrogen probes, electrical resistance probes, and corrosion spools.

Corrosion problems related to oil and gas production (drilling operation) in the presence of H_2S and CO_2 are not amenable to mitigation by dehydration or removal of corrosive species. Corrosion is controlled under these conditions by inhibition, pH control, and selection of corrosion resistant materials. Much information is available in the literature describing corrosion and sulfide stress corrosion cracking behavior of a variety of materials in sour gas service. These data can be applied to material selection and corrosion mitigation for fuel gas service.

Conclusions

Conversion to fuel gas from natural gas will require additional corrosion-control procedures. Corrosive constituents are present in fuel gas but are not found in appreciable amounts in natural gas, e.g., acid gases and ammonia, corrode common materials found in gas distribution systems. While fuel gas compositions are somewhat unique in relative amounts and mix of corrosives, experience in corrosion control in similar services is directly applicable. One of the most certain and perhaps most economical means of corrosion mitigation is to remove water from the gas prior to injection in the distribution system. Individual corrosives can also be eliminated; desulfurization is common practice. These techniques are successfully applied to the transport of coke oven gas. In specific process units, selection of more corrosion resistant materials may be necessary. An example of the latter is the elimination of high nickel alloys from units for direct burning of coke oven gas because of severe sulfidation.

It is recommended that a thorough corrosion survey of systems for materials compatibility as affected by gas conversion be made prior to any conversion, and be repeated periodically after conversion. In this way corrosion problems can be identified and suitable corrosion mitigation procedures selected.

VI. ENVIRONMENTAL CONSIDERATIONS IN RETROFIT

Emissions from the Gasification Processes

Model Steel Plant

Table 28 summarizes the major emissions from the gasification process for the model steel plant. The major points of emissions in this process are the coal storage, coal pulverizing and preparation facilities, the oxygen plant, the filter which separates water from slag and clarifier sludge, cooling tower, and the Claus sulfur recovery process.

Emissions from the coal storage pile will involve fugitive dust picked up by the wind and leachate resulting from rain water filtering through the coal pile. The coal pile should be packed tight to limit dust loss and prevent air from entering the pile causing oxidation and spontaneous combustion. Conveyors should be hooded with the hood exhaust processed through a baghouse or electrostatic precipitator. Leachate from the coal pile would resemble acid mine drainage in many respects--containing acids, organics, and soluble metals. This water should be collected and ponded for biological reduction of pollutants before being discharged to a water source.

Fugitive dust problems can be minimized by coating the coal pile with a plastic material and drawing from it only during periods of emergency. The coal normally would then be taken directly from unit train or barge by covered conveyors. The logistics of such an operation, however, would have to be carefully planned to ensure proper operation of such a system. However, care must be taken to prevent breaks in the coating which would create a chimney effect causing aspiration of air into the pile resulting in oxidation and combustion.

Emissions from the coal pulverizing preparation step consists of pollutants in the gas used in drying the coal, plus possibly some volatile constituents from the coal. A portion of the final product gas is combusted to heat air which is then supplied to the pulverizer for drying purposes. This stream is then vented from the pulverizer. The stream consists primarily of carbon dioxide, nitrogen, some water vapor, and oxygen. The stream would also contain particulates and possibly some small amounts of sulfur dioxide oxidized from the coal.

The vent stream from the coal preparation step should be processed through a baghouse or electrostatic precipitator or some other

TABLE 28. DISCHARGES FROM STEEL MILL MODEL GASIFICATION PLANT

Source	Area of Impact	Discharge		
		Flow Rate	Main Composition	Percent[(a)]
Coal storage--fugitive dust and leachate from rain	Air, Water	Dependent on wind and rain conditions	Coal dust Acids Organics Soluble Metals	--
Coal pulverizing	Air	492,160 lb/hr (or 112,144 scfm)	CO_2	2 (v)
			N_2	68 (v)
			H_2O	13 (v)
			O_2	17 (v)
			CH_4	Trace
			SO_2	Trace
			Particulates	Trace
Oxygen plant	Air	72,704 scfm	N_2	99 (v)
			CO_2,H_2O,O_2	Trace
Filter	Water, Solid Waste	21,616 lb/hr	C	17.8 (w)
			Ash	72.2 (w)
			H_2O	10.0 (w)
Cooling tower plume	Air	17,500 lb/hr	H_2O	100 (v)
			Dissolved and suspended solids	--
Claus	Solid Waste (by-product)	1,704 lb/hr	S	100 (w)
Claus tail gas	Air	675 scfm	H_2S	2.7 (v)
			CO_2	97.3 (v)

(a) (v) volume percent
(w) weight percent.

efficient particulate removal device for controlling particulate emissions. Emissions of other constituents would include some sulfur compounds such as SO_2, though these emissions should be relatively small. Also, in any coal crushing operation, considerable noise is generated and the pulverizing operation should be housed in a building to minimize this effect.

To limit dust loss, the entire coal pulverization facility should be located in a building with positive ventilation control. The exhaust from the building would then be processed through a particulate control device.

The discharge from the oxygen plant would involve primarily nitrogen which is not considered a harmful emission and would require no control.

Wet slag from the gasifier along with clarifier sludge from the water scrubbing operation is processed through a filtration step for liquid/solids separation. The slag from the gasifier contains a variety of constituents typical of coal ash but due to the high temperature in the gasifier is relatively inert and not expected to be a pollution problem; however, actual operating data will be necessary to verify this. Sludge from the clarifier, however, would contain dissolved gases such as H_2S which could present an odor problem. Lime could be added to the clarifier circuit to fix the H_2S in a nonvolatile form, or with highly alkaline coals, the alkalinity in the slag from the gasifier may be sufficient to alleviate the problem(16).

A significant discharge to the atmosphere would be the cooling tower plume. The cooling tower water would contain dissolved constitutents from the scrubber circuit that overflows from the clarifier. These constituents would be present to some extent in the drift loss or plume from the cooling tower. Although many of these compounds may be present only in infinitesimal amounts when combined with the water in the plume, they may create a corrosion or health menace in the area around the plant. A solution to this problem is to use dry cooling towers or a cooling pond either of which would involve much greater cooling area.

About 772 kg/hr (1704 lb/hr) of elemental sulfur would be produced from the Claus plant in this process. This sulfur would be of marketable quality and could be stored and shipped. The Claus process, however, only removes about 95 percent of the sulfur compounds of the inlet stream. The resulting tail gas or vent stream from the Claus process, therefore, would contain hydrogen sulfide and CO_2. With the system shown and if meeting regulations with the least direct cost were an objective, then this tail gas could be blended with the product gas from the gasifier and combusted without exceeding even the strictest state limitations on sulfur dioxide emissions.

Model Refinery Plant

The major emissions from the refinery model gasification plant are shown in Table 29. Sources of emissions are coal storage, the gasifier itself, scrubber effluent, and emissions from the Stretford desulfurization process.

Effluents from coal storage would involve similar considerations to those discussed for the steel plant model. Because crushed, sized coal would be purchased from the mine, however, dust loss for the refinery would be less than for the steel plant due to the lower percentage of fines or small particles. Also, air and noise pollution from drying and crushing operations in the coal preparation step would not be present. If these operations were installed, similar consideration to those for the steel plant model would have to be employed.

About 801 kg/hr (1768 lb/hr) of dry ash would be emitted from the gasifier in the form of bottom ash. The Wellman-Galusha is a "dry ash" or nonslagging process and the bottom ash may have characteristics similar to that from a stoker or pulverized fired boiler. Common practice in boiler installations is to truck or sluice the ash to pond or landfill.

The effluent from the scrubber system contains significant amounts of tars, ammonia, and phenols, which would have to be treated prior to disposal. In some cases these products may be able to be used in the industrial plants or marketed. For instance, in the case of the refinery,

TABLE 29. DISCHARGES FROM REFINERY MODEL GASIFICATION PLANT

Source	Area of Impact	Discharge		
		Flow Rate	Main Composition	Percent
Coal storage--fugitive dust and leachate from rain	Solid waste	Depends on wind and rain conditions	Coal dust Acids Organics Soluble metals	--
Scrubber effluents	Water			
Tar separation		1153 lb/hr	Tar	91
			H_2O	9
NH_3 stripping		1095 lb/hr	NH_3	20
			H_2O	80
Phenols		120 lb/hr	Phenols	100
Cyanide		--	HCN	--
Hydrocarbons		--	CxHy	--
Pond	Air			
Evaporation		Trace	Ammonia	Trace
			Phenols	"
			Hydrocarbons	"
			HCN	"
Stretford	Solid waste (by-product)	777 lb/hr	Sulfur	100
			Sodium	
			Thiosulfate & sodium Thiocyanate	--

the recovered tars could possibly be used to supplement residual oil in making of asphalt. However, this would have to be evaluated as to the effect of these tars on the asphalt production process of the plant. Tars would be recovered by decantation and would result in a composition of about 91 percent tar, and 9 percent water. Ammonia and other compounds, such as trace amounts of hydrogen sulfide which may be dissolved in the scrubber water, could be steam stripped and recovered for sale. Phenols could also be recovered for use by use of the Phenolsolvan process, or they could be biologically reduced to sludge and separated from the water for disposal. There is no immediate use for phenols in the refinery so biological reduction would probably be employed. The economics of this versus recovery of the phenols in a potentially more expensive process would have to be evaluated further.

In addition to tars, ammonia, and phenols, the scrubbing water could also contain small amounts of hydrogen cyanide (HCN) and hydrogen fluoride (HF). Hydrogen cyanide in the water stream can be very detrimental to a biological control process, and it may have to be treated separately. Otherwise, it would be expected to follow hydrogen sulfide through the process. Hydrogen fluoride would react with the ash in the coal and be disposed of in a neutralized form with the ash.

Because many of the constituents in the scrubbing water are highly volatile and odorous, care must be taken throughout the water scrubbing and treatment system to minimize leaks and evaporation. Reaction vessels should be covered and vented either back to the scrubbers or to some other control process. Also, if the volatile and odorous constituents are not removed from the water before being discharged to the settling pond, odors could result from pond evaporation.

The recovery or disposal of tars, ammonia, phenols, and other gas liquor constituents will involve some hydrocarbon emissions. These emissions result from leaks around seals in pumps and storage facilities. Refineries, in general, are accustomed to dealing with the problems associated with handling these compounds, however, and should be able to handle the additional load supplied by a coal gasifier.

The Stretford desulfurization process for this particular design would produce 349 kg/hr (770 lb/hr) of elemental sulfur which could be stored and sold. The Stretford purge stream will contain sodium salts of anthraquinone disulfonate, metavanadate, citrate, thiosulfate, and thiocyanate. This stream may require special treatment or disposal methods (40).

Emissions from Combustion Processes

Emissions from combustion processes result generally from four types of pollutants; emissions of sulfur dioxide, oxides of nitrogen, particulates, and trace constituents, such as polycyclic organic matter or heavy metals.

Emissions of Sulfur Dioxide

In the gasification process many sulfur compounds in the coal are converted to sulfur compounds in the gas. The major sulfur-bearing constituent is hydrogen sulfide with minor amounts of carbonyl-sulfide (COS), carbon disulfide (CS_2), and mercaptans. If these compounds are not removed from the fuel gas prior to combustion, they are oxidized quantitatively to sulfur dioxide in the products of combustion. These expected emissions, if all the sulfur in the coal is converted to sulfur in the gas, are shown in Figure 33 as a function of coal sulfur and heat content.

Standards have not yet been developed specifically for dealing with sulfur emissions from coal gasification applications as described in this study. There is currently debate on whether sources fired with gas from coal should be treated as solid fuel fired or gas fired sources and as to whether emissions should be based on the heating value of the gas or solid fuel.

If emissions are based on the heat content of the coal, then they are a function of coal sulfur and heat content as shown in Figure 33. As can be seen from Figure 33, a coal-sulfur content of less than 0.5 to 0.8 percent

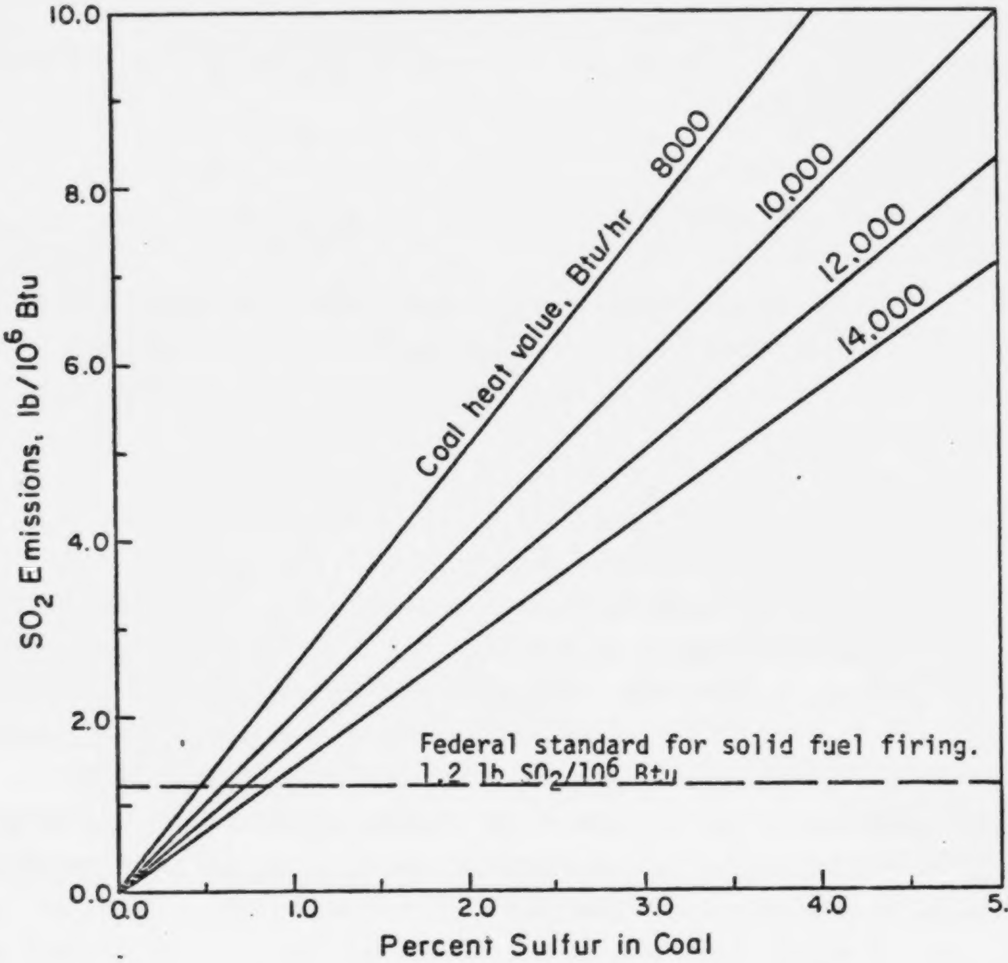


FIGURE 33. SO_2 EMISSIONS VERSUS SULFUR IN COAL

would be required for most coals before compliance with the Federal New Source Performance Standard for solid fuel fired sources of 2.16 kg $SO_2/10^6$ kcal (1.2 lb $SO_2/10^6$ Btu) heat input could be met without some form of desulfurization.

Figure 34 shows the expected emissions of sulfur dioxide from combustion processes based on the heat and sulfur content in the fuel gas. As can be seen, to meet the Federal standard for new sources based on solid fuel firing, a sulfur level in the fuel gas of about 1000 ppm would be allowable for low-energy gas with a heating value of 5.59 kJ/Nm^3 (150 Btu/scf). As the heating value of the gas increases, the allowable sulfur content also increases.

Many states, however, have tighter standards for SO_2 emissions and it appears that the trend is for tighter standards to be promulgated. New Mexico has established one of the strictest standards for SO_2 emissions from solid fuel-fired sources -- 0.61 kg $SO_2/10^6$ Kcal input (0.34 lbs of $SO_2/10^6$ Btu input). Meeting this standard would limit the sulfur concentration in fuel gas with a heating value of 5.91 MJ/Nm^3 (150 Btu/scf) to 300 ppm or less. However, New Mexico has proposed a much stricter standard of 0.07 kg $SO_2/10^6$ Kcal (0.04 $lb/10^6$ Btu) for gasification plants involved in producing SNG. Whether this standard would also apply to gasification plants producing lower heating value fuel gases is uncertain.

In addition to environmental limitations on sulfur content in the fuel gas, there are also certain process considerations in an industrial application. Hydrogen sulfide is known to have a high corrosion potential in piping and distribution systems, especially when in the presence of water vapor or oxygen. Also, when firing the gas directly in a furnace, sulfur compounds in the fuel gas (such as hydrogen sulfide) can cause problems with sulfidation of certain kinds of products, particularly high-grade steel products. Determining the maximum limit of sulfur compounds in the gas to prevent these problems from occurring will require further definition; however, it is possible that these requirements may be more restrictive than environmental requirements in some cases. A more complete discussion of the potential deleterious effects of fuel gas contaminants on distribution systems and products is given in Section V.

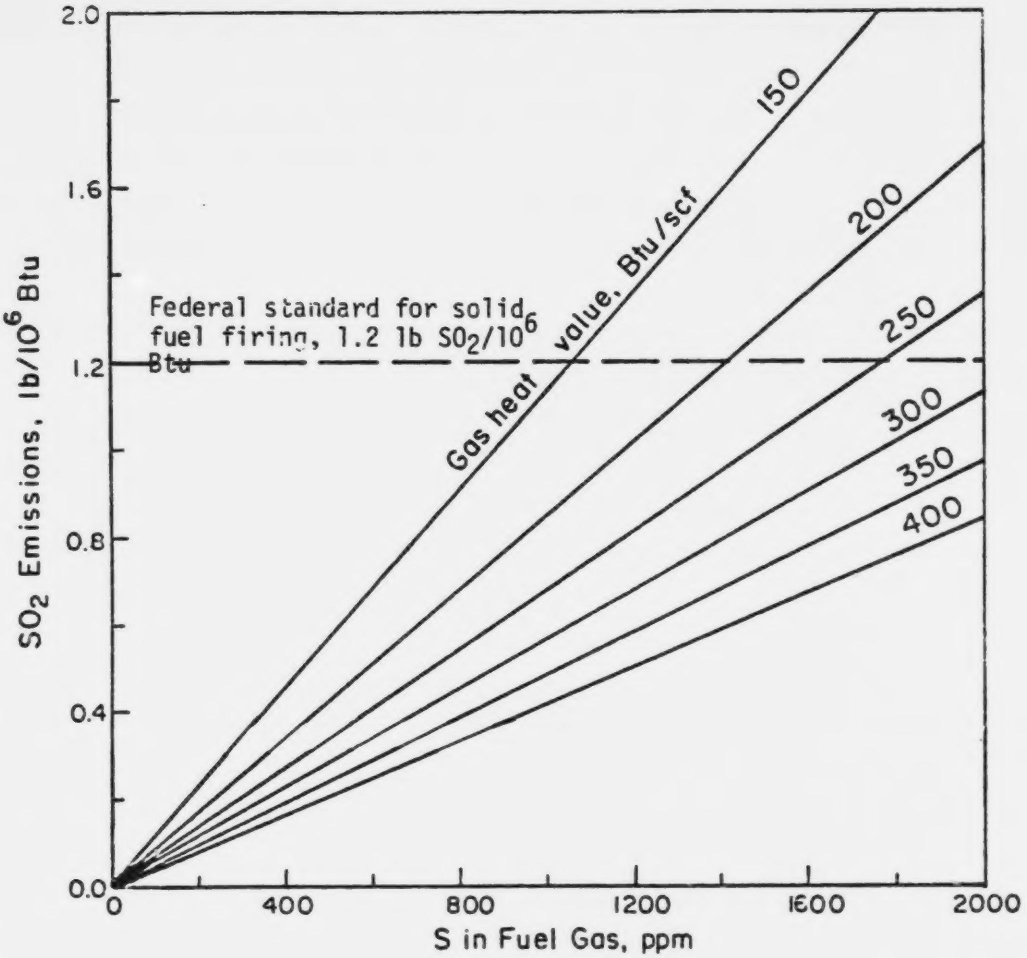


FIGURE 34. SO_2 EMISSIONS VERSUS SULFUR IN FUEL GAS

For this study a maximum sulfur concentration in the fuel gas of 300 ppm was established. This limitation would allow meeting the Federal standard for solid fuel firing of 2.16 kg $SO_2/10^6$ Kcal (1.2 lbs $SO_2/10^6$ Btu) based on the fuel gas heating value for both cases described in this study. Additionally, it is also a reasonable lower bound on what can easily be achieved by atmospheric pressure, chemical-absorption type sulfur removal systems, such as those analyzed in this study, without unusually high operating cost or complex sulfur recovery processes.

The expected emissions of sulfur compounds for both hypothetical industry plants studied are given in Table 30. Emissions are given as a function of both the heat content of gas fired and the heat content of coal gasified. In the case of the steel plant model, the disposition of the Claus plant tail gas must be considered. The tail gas, consisting primarily of CO_2 and H_2S, could be handled in several different ways.

(1) The tail gas could be processed through a Stretford or other type of liquid phase oxidation system to remove the H_2S and convert it to elemental sulfur.

(2) The tail gas could be combined with the clean gas and burned in the plant processes.

(3) The tail gas could be incinerated or burned in a boiler.

TABLE 30. EXPECTED EMISSIONS OF SULFUR DIOXIDE FROM COMBUSTION PROCESSES IN MODEL PLANTS

		Emissions, kg $SO_2/10^6$ Kcal ($lb SO_2/10^6$ Btu)		
	kg SO_2/day (lb SO_2/day)	Based on Gas Energy Burned in Processes	Based on Coal Energy Gasified	
Steel Plant Model	1778 (3925)	0.316 (0.176)	0.222 (0.124)	clean gas only
	3869 (8541)	0.688 (0.383)	0.480 (0.267)	clean gas with Claus tail gas combined
Refinery Model	2020 (4460)	0.359 (0.200)(a)	0.417 (0.232)	

(a) Low-energy gas plus refinery waste gas.

Table 30 shows expected sulfur emissions for Cases (1) and (2) above. Case (1) would be considered an expensive solution but would minimize total atmospheric sulfur emissions. The complexity of the plant would increase along with the amount of elemental sulfur that would have to be handled. Case (2) represents the simplest solution but results in nearly double the total atmospheric sulfur emissions. Case (3) would result in the same total emissions as Case (2) if the tail gas were incinerated with no sulfur controls. If the tail gas were burned in a coal-fired boiler, which might be used for raising steam for operating the gasification plant, SO_2 scrubbers could be used on the boiler to reduce the overall sulfur emissions.

Emissions of Oxides of Nitrogen

This discussion is to evaluate the probable change in NO_x emissions that would result when changing from the combustion of natural gas to the combustion of one of the moderate or low heating value fuels considered in this study. The case in which there is no fuel-bound nitrogen will be considered first. Then the effect of fuel-bound nitrogen, specifically in the form of ammonia, will be considered.

Figure 35 shows the equilibrium nitric oxide concentration as a function of the percent theoretical air for several different air preheats of natural gas-air mixtures. The rapid increase of NO with air up to about 25 percent excess air, followed by a fall-off, is obvious from Figure 35. Figure 36 shows, however, that for constant combustion temperatures, the NO concentration tends to level off at a constant value as the percent theoretical air increases. It is clear, then, that the equilibrium NO concentration increases with increase in available oxygen and with combustion temperature. Gas composition has little effect on the curves of Figure 36 if these two factors are used as basic values. The largest change is to adjust the NO concentration linearly with the N_2 concentration in the stoichiometric mixture(42).

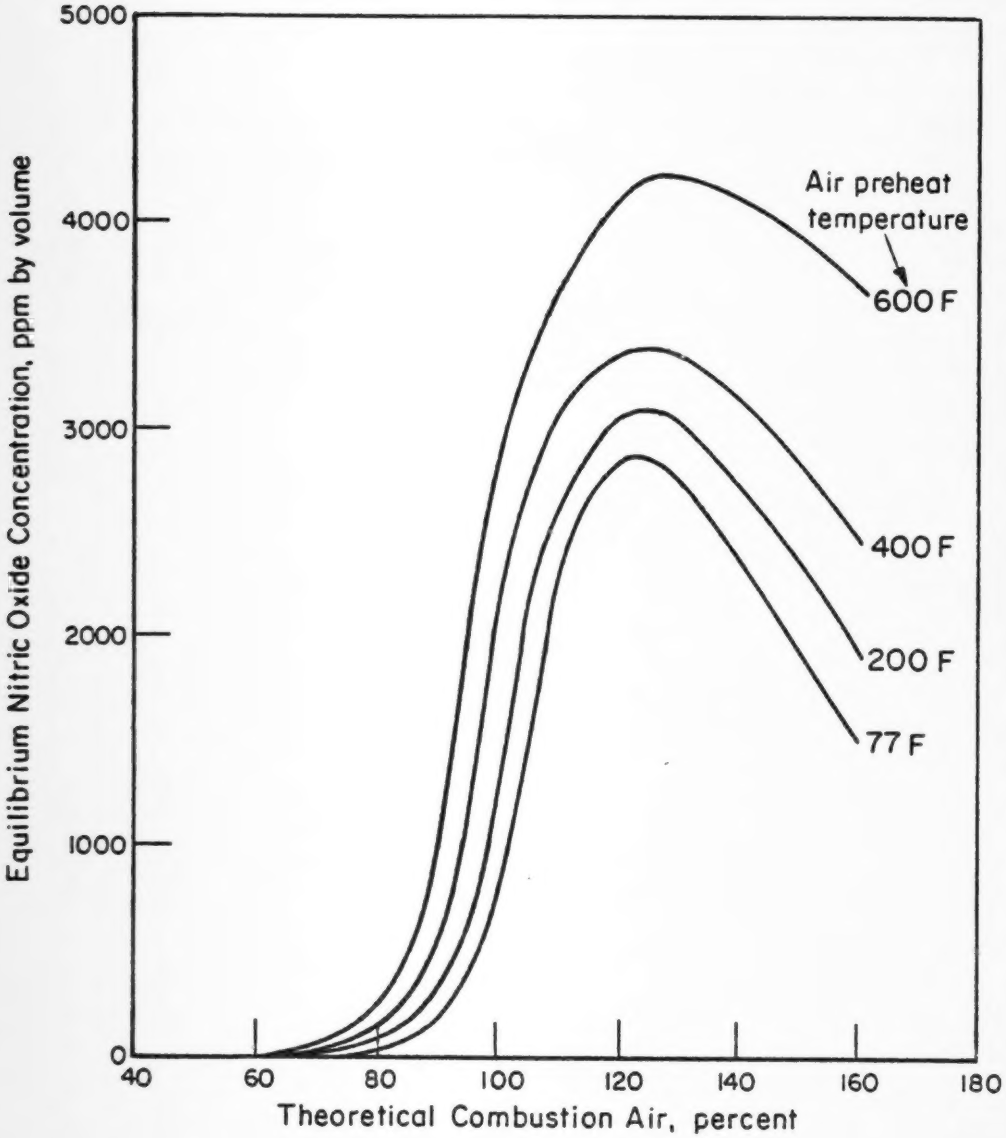


FIGURE 35. EFFECT OF AIR PREHEAT ON NITRIC OXIDE EQUILIBRIUM(40)

Thermodynamic equilibrium data for natural gas fuel

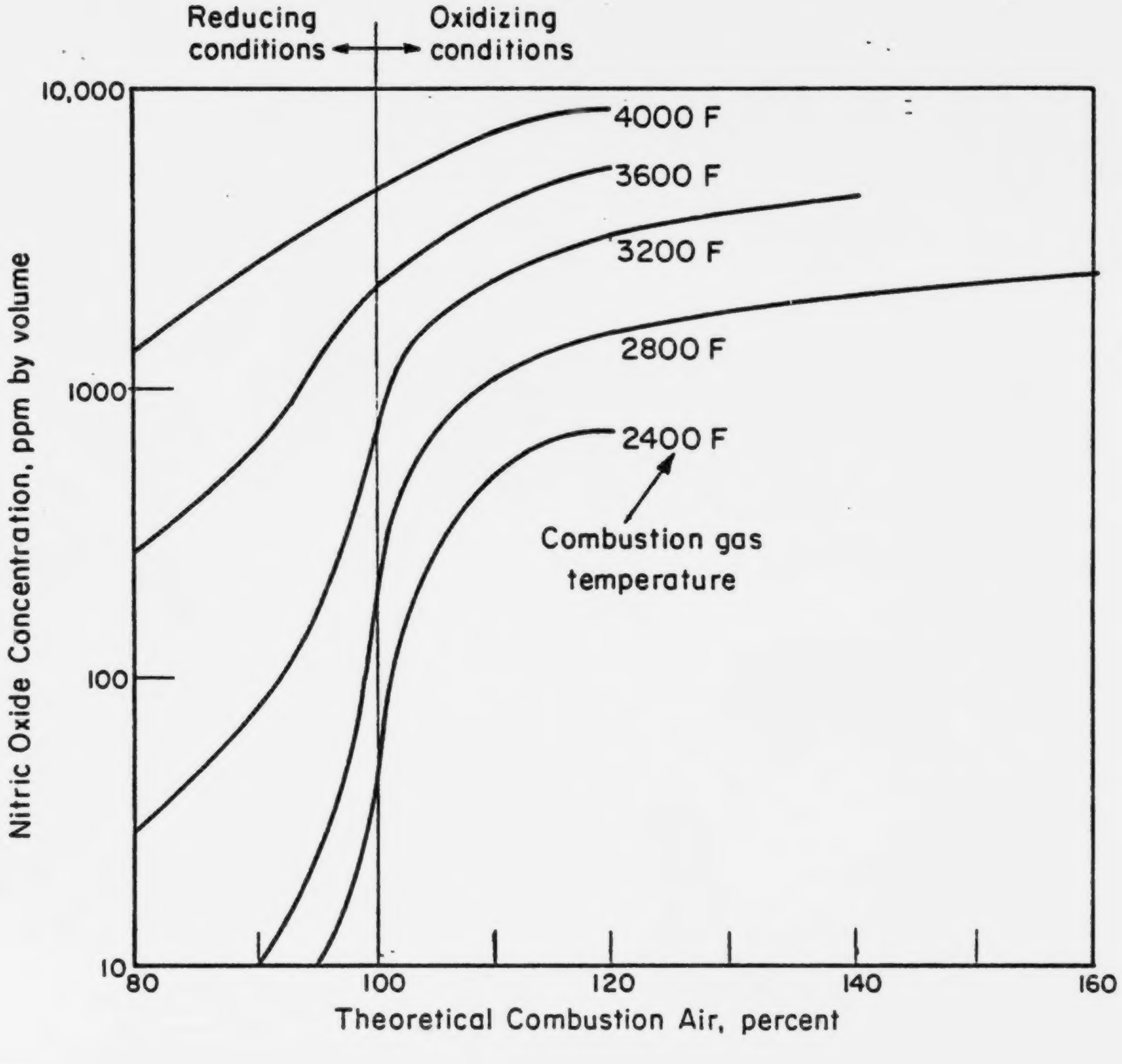


FIGURE 36. EFFECT OF TOTAL AIR ON NITRIC OXIDE EQUILIBRIUM(40)
Thermodynamic equilibrium data for natural gas fuel

Thus, in comparing natural gas with the other fuels of concern in this study, and assuming a constant percent theoretical air, the adiabatic flame temperature and amount of N_2 in the stoichiometric mixture are the primary considerations for the case with no fuel-bound nitrogen. This permits the specific heating value of the fuel, the stoichiometric fuel/air ratio, and the air and/or fuel preheat to be neglected as considerations.

The next factor to consider is the effect of the available reaction time. This is significant because the rate of production of thermal NO is slow compared to the combustion times available or needed in most furnaces. This is why a maximum value of 175 ppm of NO when burning natural gas with 15 percent excess air is reasonable, whereas the equilibrium value is about 3000 ppm (Figure 35). Figure 37 shows the effect of residence time on curves comparable with Figure 36. These values can be compared with current New Source Emission Standards for large boilers of 175 ppm, 230 ppm, and 575 ppm of NO_x for 15 percent excess air burning gas, oil, and coal, respectively. These correspond to 0.36, 0.54, and 1.26 kg NO/10^6 kcal (0.2, 0.3, and 0.7 pounds of NO per 10^6 Btu). It is seen from Figure 37 that at a combustion gas temperature of 1760 C (3200 F), the NO concentration only reaches 1/8 the equilibrium concentration shown in Figure 36 in 0.4 sec. (At only 3.05 m/sec, this would be a distance of 1.22 m.) For 2000 C (3600 F), the ratio is about 1/10. One may conclude then that for flames at the same firing rate, same temperature, and same excess air, there will be little difference in the actual concentration of thermal NO.

A computation can now be made of the relative NO_x values for different fuels operating under the same excess air conditions and same initial temperature, providing no fuel-bound nitrogen (discussed below) is present. Four fuels are considered, a natural gas (Table A-1), a Koppers-Totzek gas considered as a replacement for natural gas in a secondary steel plant, and a Wellman-Galusha gas mixed in proposed winter and summer proportions with a refinery gas. The equilibrium NO at the adiabatic flame temperature with 10 percent excess air is computed for each of these gases from Figure 36, correcting the concentration value by the ratio of N_2 in the raw mixture to that

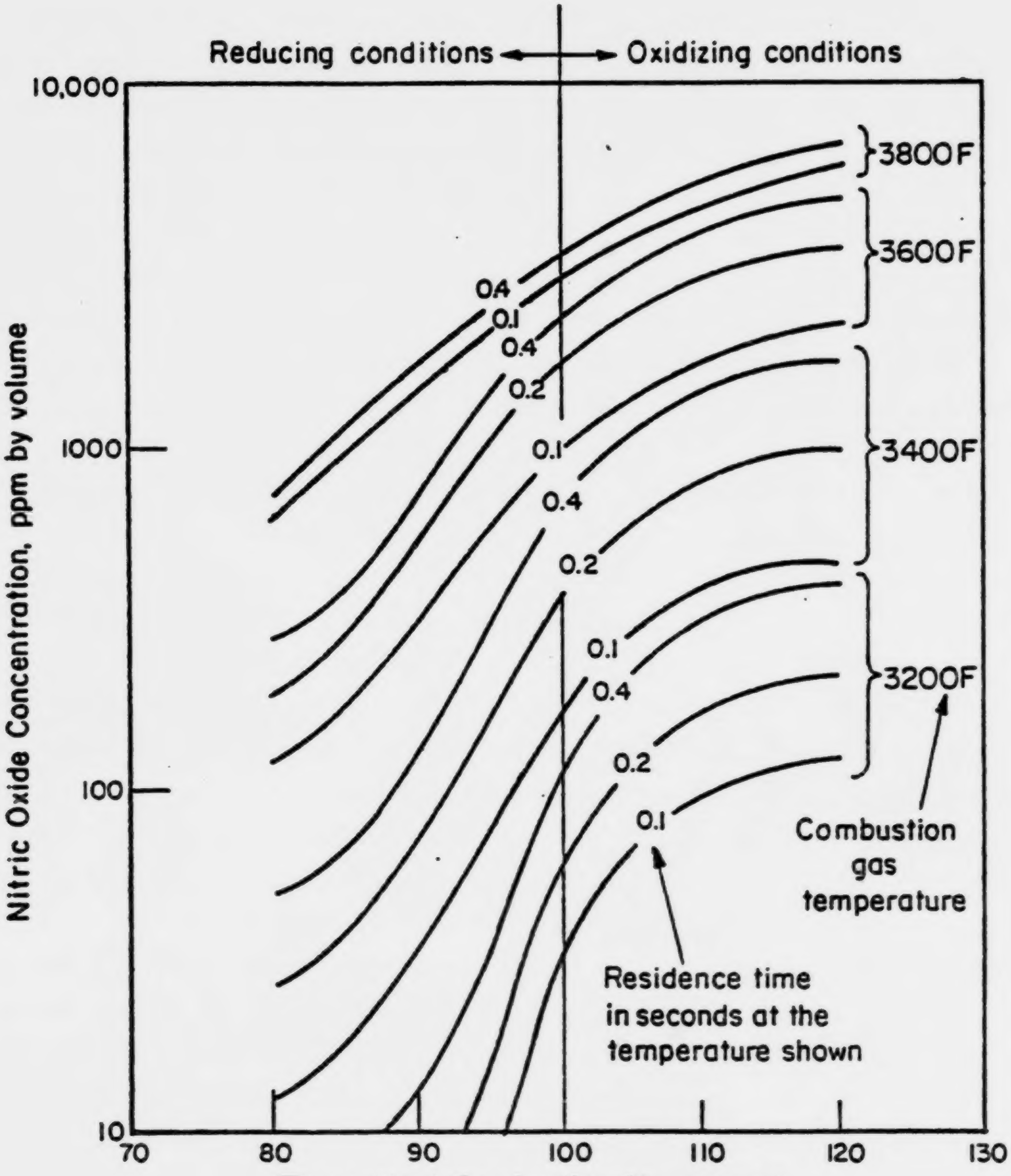


FIGURE 37. EFFECT OF TOTAL AIR, FLAME TEMPERATURE AND RESIDENCE TIME ON NITRIC OXIDE CONCENTRATIONS(40)

Kinetic data for natural gas fuel

for natural gas. The value is then further corrected for the heating value of the raw mixture*. It is seen from Table 31 that the low temperature of the refinery gas/Wellman-Galusha gas mixture reduces the NO production the most, in spite of the high nitrogen content of the fuel. The high heating value of the Koppers-Totzek gas per unit mass of products, plus the increased volume ratio of fuel-to-air, more than compensates for the increased temperature of the Koppers-Totzek gas and, thus, also results in lower NO production. As a result, if essentially all fuel-bound nitrogen is stripped from the moderate and low-heating value gases, these gases will give lower NO output than natural gas under similar firing conditions.

TABLE 31. RELATIVE NO PRODUCTION FROM THERMAL FIXATION OF VARIOUS FUELS AT 10 PERCENT EXCESS AIR

Fuel	Relative NO_x Production
Natural gas	1.00
Koppers-Totzek	0.93
Wellman-Galusha, winter mix with refinery gas	0.72
Wellman-Galusha, summer mix with refinery gas	0.74

Nitrogen bound in various fuel constituents (primarily NH_3 for gas from coal) does not convert to NO_x by the same process as thermal fixation of elemental nitrogen. In the case of fuel-bound nitrogen, temperature and time are of little importance. The two major factors are stoichiometry (or percent excess air) which determines the amount of oxidant available, and the concentration of nitrogen compounds in the fuel.

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*As an example, the Koppers-Totzek gas may be compared with natural gas. The adiabatic flame temperatures with 10 percent excess air are 2190 K and 2140 K. From a cross-plot of Figure 36, the concentration of NO is 3400 and 3000 ppm, respectively. The raw mixtures have 63.0 and 75.3 percent N_2, respectively; using the ratio, the value of 3400 is corrected to 2845. The heating value of the raw mixtures are 1098 and 1074 Btu/lb, respectively; correcting to a common heat input, 2845 ppm of NO becomes 2783 ppm of NO. The ratio of this value to 3000, which is 0.928, is the NO_x production of the Koppers-Totzek mixture relative to natural gas.

The effect of stoichiometry on conversion of NH_3 to NO_x in a methane flame is shown in Figure 38. These data are for a premixed flame where fuel and oxidant are thoroughly mixed prior to burning. In nozzle mix-type burners, which are more common in industry than premix burners, the stoichiometry in the flame is primarily a function of the mixing rate between the fuel gas and combustion air. In these cases the amount of combustion air would have only a minor effect on NO_x conversion.

The major effect of fuel-bound nitrogen conversion to NO_x in nozzle mix-type burners is nitrogen concentration in the fuel. Turner, at al.,(43) have shown that the form in which the nitrogen is bound in the fuel has no effect on nitrogen conversion.* Figure 39 shows the conversion of fuel-bound nitrogen to NO_x in a Rankine-cycle can-type combustor using a liquid fuel with pyridine as the nitrogen carrying additive. The data for the curve in Figure 39 represent a wide range of excess airs of from 120 to 175 percent of theoretical air and indicate that, within this range of excess air levels, excess air has little or no effect on nitrogen conversion.

Figure 40 shows a compilation by Dykema and Hall(46) of utility boiler data over a wide range of nitrogen concentrations of the mass fraction of fuel-bound nitrogen converted to NO_x. The curve of Figure 38 and corresponding data of Hazard (Figure 39) are added. While the scatter of the data is large, it must be realized the there is an arbitraryness in accounting for the amount of thermal nitrogen to be deducted from the total; the higher fraction mass conversion calculation is particularly sensitive to this effect. Nevertheless, the trend of the data is obvious.

Data on fuel-bound nitrogen, which predominantly consists of NH_3, in fuel gas from coal is limited. Table 32 illustrates some typical ammonia concentrations in raw uncleaned gas from various types of coal gasification processes. In cold gas desulfurization processes, much of this ammonia (at least 90 percent) would be removed in the water scrubbing step preceding desulfurization, as long as the scrubbing water is continuously stripped of absorbed ammonia. Additional ammonia may be removed in the desulfurization step depending on the sorbents used.

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*This analysis involved liquid fuels with eight different nitrogen containing additives.

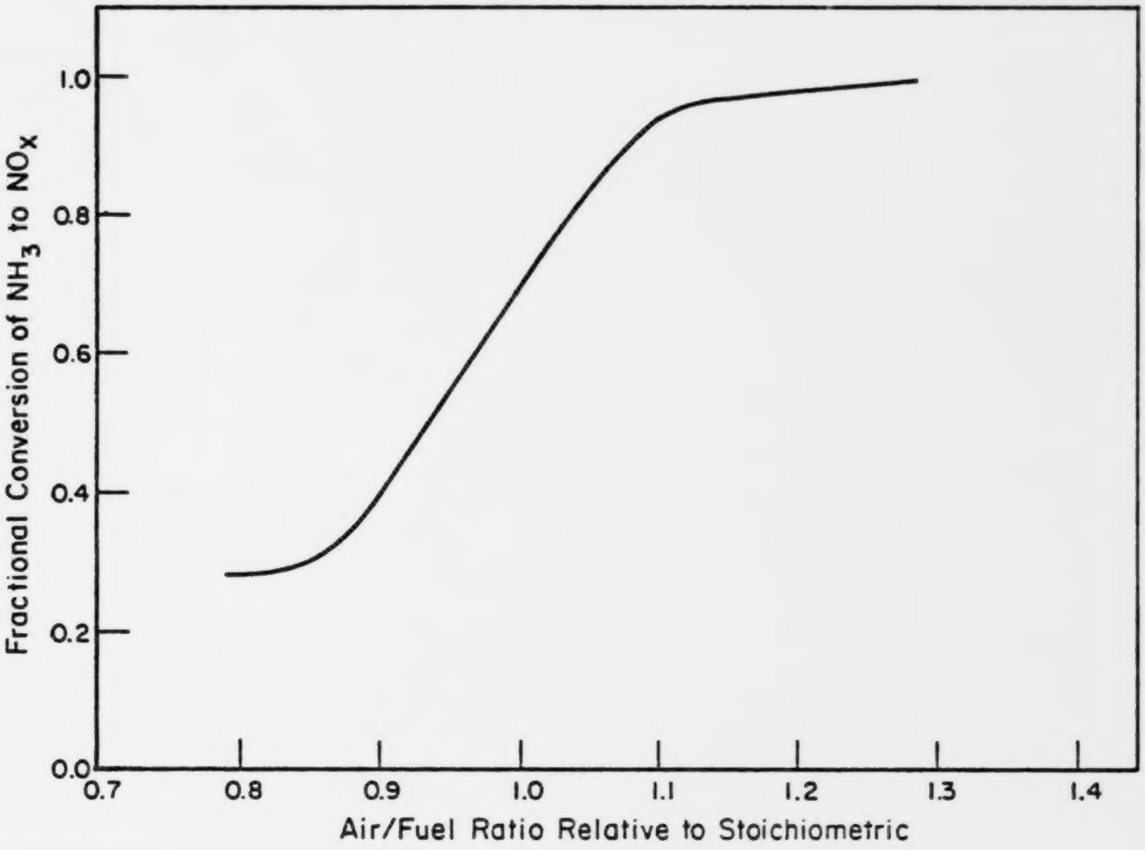

FIGURE 38. FRACTIONAL CONVERSION OF NH_3 IN PREMIXED METHANE-AIR MIXTURE*(14)

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* NH_3 equivalent to 1200 ppm NO and air consisting of oxygen-helium mixture.

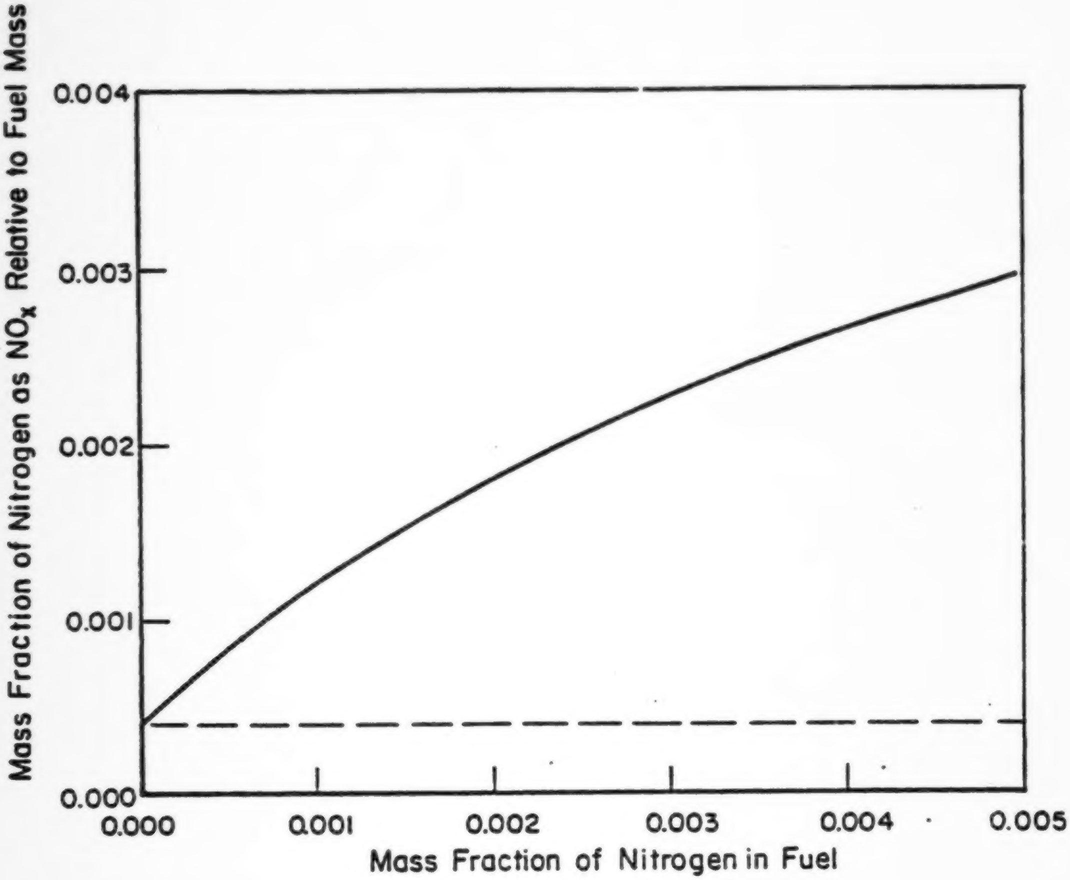

FIGURE 39. FUEL NITROGEN IN LIQUID FUEL-FIRED RANKINE-CYCLE COMBUSTOR CONVERTED TO NO_x*(44)

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*Using pyridine as the nitrogen source and using ASTM Jet A combustor. Tests covered from 120 to 175 percent theoretical air.

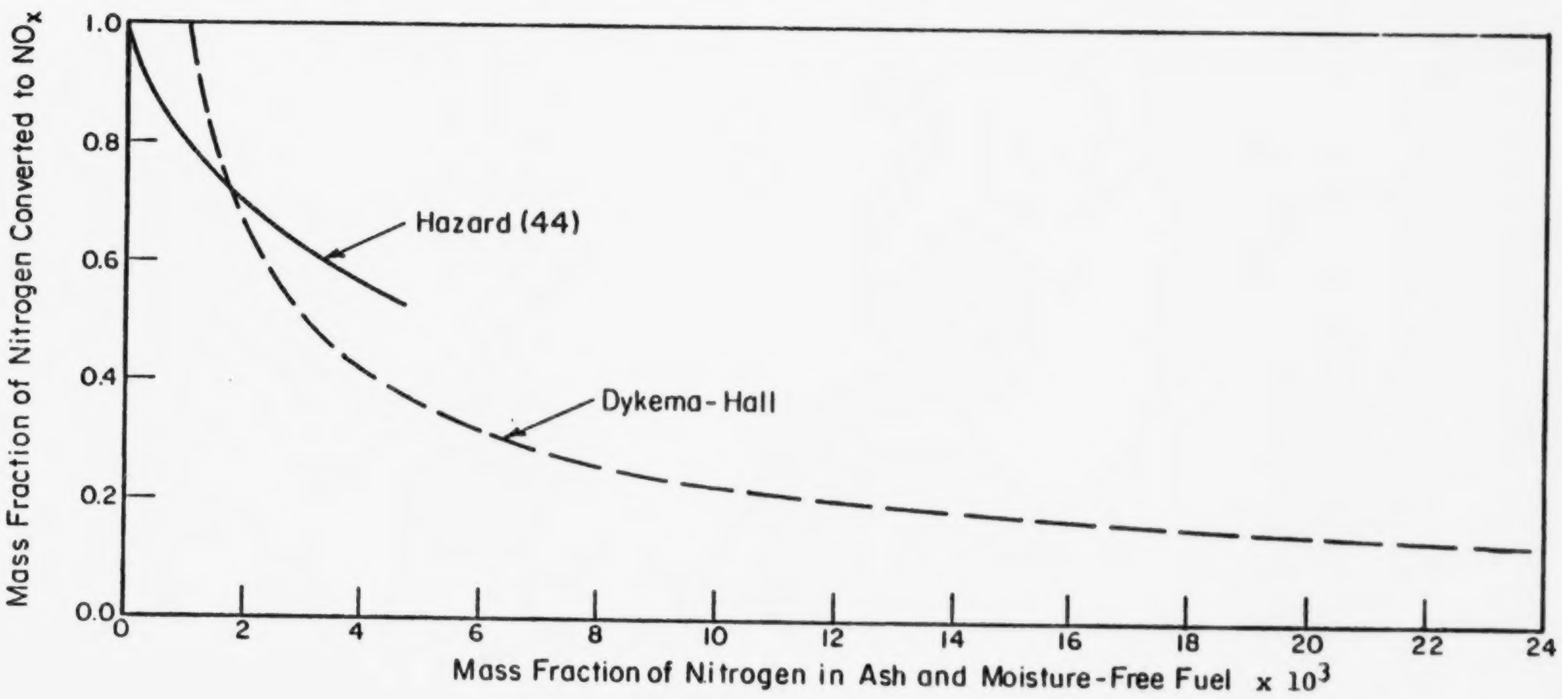

FIGURE 40. FUEL NITROGEN CONVERTED TO NO_x (44,45)

TABLE 32. TYPICAL AMMONIA CONCENTRATIONS IN RAW UNCLEANED FUEL GAS FROM COAL

Gasifier	Ammonia Volume Percent	Ammonia lb/lb fuel x 10^3	MJ/Nm^3	(Btu/scf)	Reference
Koppers-Totzek					
single stage entrained slagging (O_2 blown)	0.17	1.13	11.3	(286)	46
two-stage entrained slagging (air blown)	0.38	2.53	4.9	(125)	47
Lurgi					
pressurized fixed-bed (O_2 blown)	0.70	4.66	12.7	(323)	48
atmospheric fixed bed (air blown)	0.25	1.46	5.5	(139)	47

Table 33 shows expected emissions of NO (in lb/10^6 Btu of heat input) for the ammonia concentrations shown in Table 32 using the curve of Dykema and Hall from Figure 40. Values are given for both the raw gas and assuming 90 percent ammonia removal. In addition, Table 33 gives estimated emissions due to thermal fixation of N_2 assuming a thermal contribution of 100 ppm NO in the flue gas on stoichiometric mixture. Total expected NO emissions from both thermal fixation and oxidation of fuel-bound nitrogen assuming 90 percent NH_3 removal are also given.

TABLE 33. ESTIMATED EMISSIONS FROM
RAW AND CLEANED FUEL GASES

Gasifier	NO Emissions, Kg/10^6 Kcal (lb/10^6 Btu): Due to NH_3 in Gas: Raw Gas	NO Emissions, Kg/10^6 Kcal (lb/10^6 Btu): Due to NH_3 in Gas: Assuming 90% Removal of NH_3	NO Emissions, Kg/10^6 Kcal (lb/10^6 Btu): Thermal NO Assuming 100 ppm NO in Stoichiometric Mixture	NO Emissions, Kg/10^6 Kcal (lb/10^6 Btu): Total NO With 90 Percent NH_3 Removal
Koppers-Totzek				
single-stage entrained slagging	0.79 (0.44)	0.08 (0.047)	0.15 (0.086)	0.24 (0.133)
two-stage entrained slagging	2.63 (1.46)	0.47 (0.259)	0.22 (0.120)	0.68 (0.379)
Lurgi pressurized fixed bed	1.13 (0.63)	0.31 (0.171)	0.16 (0.092)	0.47 (0.263)
Atmospheric fixed bed	2.09 (1.16)	0.25 (0.141)	0.20 (0.112)	0.45 (0.253)

For the raw gas expected emissions of NO from oxidation of fuel-bound nitrogen alone would exceed the Federal standard of 1.26 Kg NO/10^6 kcal (0.7 lb NO/10^6 Btu) for coal-fired systems in two cases and would exceed the standard for gas-fired systems of 0.36 Kg NO/10^6 kcal (0.2 lb NO/10^6 Btu) in all cases. With 90 percent removal of NH_3 the expected NO emissions including the assumed contribution from thermal fixation would be less than the coal standard of 1.26 Kg NO/10^6 kcal (0.7 lb NO/10^6 Btu) in all cases and would approach the gas standard in most cases.

In both industry systems considered in this study, NH_3 is assumed to be entirely removed in the combination of water scrubbing and amine or Stretford desulfurization. Emissions of NO would, therefore, consist primarily of those from thermal fixation of elemental nitrogen. Under these circumstances NO emissions overall would decrease relative to those with natural gas as was shown in Table 31.

Particulate Emissions

Particulate content in the final clean product gas from both of the gasification plant models is negligible. Combustion of this gas, therefore, would be expected to result in negligible particulate emissions to the atmosphere and no particulate control would be required. In both model plant cases the low-energy gas would be replacing the firing of some heavy oil which would result in an overall decrease in particulate emissions from these two industries.

Emissions of Trace Constitutents

Emissions of trace organic constituents such as polycyclic organic matter (POM) are a function of the number of long chain hydrocarbons or ring-type-hydrocarbons in the fuel itself and of the combustion conditions. Coal and oil both contain significant quantities of these compounds. However, the product gas from gasification, should contain few, if any, long chain or ring-type hydrocarbon components. Combustion conditions for firing the fuel gas would be similar to those for firing natural gas. Thus, emissions of these types of materials would be expected to be similar to that of natural gas. They would be significantly less than if the coal were fired directly or if oil were used directly as the fuel.

Other trace constituents, such as trace metals that may be vaporized in the combustion process, are also potential pollutants. The more volatile metals (mercury, etc.) would be vaporized in the gasification process but should be condensed in the water scrubber and cooling sections of the gas-cleaning processes. The ultimate fate of these constituents must still be determined in order to assess the true environmental impact.

VII. POTENTIAL IMPACT OF ADVANCED HOT GAS CLEANING SYSTEMS

All fuel gas desulfurization systems that are applicable to cleaning gas from coal and have been proven commercially successful are at gas temperatures of less than 250 F. The two processes used in this study, the MDEA and Stretford systems, operate at temperatures of ambient or slightly above. The raw fuel gas from a gasifier, however, contains significant amounts of sensible heat which could represent from 10 to 20 percent of the energy in the raw gas, depending on the process and the raw gas temperature. There has been considerable emphasis on developing fuel gas desulfurization processes capable of cleaning fuel gas at elevated temperatures. This would allow the gas to be fired hot, thus, conserving the sensible heat and increasing the overall thermal efficiency. This concept has obvious merit, especially for power plant applications where the hot gas needs only to be piped a short distance to the point of combustion. However, different considerations are necessary for industrial plants. Therefore, an evaluation was made of the relative advantages and disadvantages such systems might have in an industrial situation.

Table 34 lists the leading hot-gas desulfurization systems under development. These processes can generally be classed as those using fully calcined dolomite of half-calcined dolomite (Consolidation Coal and Air Products and Chemicals), those using iron oxide (Bureau of Mines and Babcock & Wilcox), and those using molten salt baths (Battelle-Northwest).

The dolomite processes operate at the highest temperatures [from about 815 to 1100 C (1500 to 2000 F)] and regeneration yields an H_2S-rich gas suitable as a Claus feed. Regeneration of these processes is accomplished with steam and CO_2 acording to the following reaction:

$$CaS \cdot MgO + H_2O + CO_2 \rightarrow CaCO_3 \cdot MgO + H_2S.$$

This reaction is for the Consolidation Coal half-calcined dolomite process. The Air Products and Chemicals full-calcined dolomite process has been abandoned due to poor sorbent regenerability(50).

TABLE 34. ADVANCED HIGH-TEMPERATURE CLEANING SYSTEMS UNDER DEVELOPMENT

Process	Sorbent	Temperature, C (F)	Sulfur Form	Status
Consolidation Coal*	$CaCO_3 \cdot MgO$	816-982 (1500-1800)	H_2S	Pilot
Air Products and Chemicals	$CaO \cdot MgO$	871-1093(1600-2000)	H_2S	Abandoned
Bureau of Mines	Fe_2O_3 + fly ash	423-816 (800-1500)	SO_2	Pilot
Babcock & Wilcox	Fe_2O_3	371-649 (700-1200)	SO_2	Experimental
Battelle-Northwest	$NaCO_3 + CaCO_3$	593-923(1100-1700)	H_2S	Pilot

*Conoco Coal Development Corporation

Processes using iron oxide as a sulfur sorbent operate at temperatures of about 370 to 815 C (700 to 1500 F). The sorbent is regenerated with air yielding an SO_2-rich gas stream by the following reaction:

$$2FeS + 3\text{-}1/2\ O_2 \rightarrow 2SO_2 + Fe_2O_3.$$

The SO_2 can then be reduced to elemental sulfur, converted to sulfuric acid, or converted to $CaSO_4$ with lime or limestone scrubbing.

The molten salt process operates at temperatures of from 593 to 923 C (1100 to 1700 F) and absorb sulfur compounds in a molten solution of $NaCO_3$ and $CaCO_3$. The sorbent is regenerated with steam and CO_2 yielding an H_2S rich gas stream suitable for feed to a Claus sulfur recovery unit.

At the present time, none of the hot gas cleanup systems discussed are commercially available. At present, all are in the pilot stage of development with the exception of Babcock & Wilcox, which is experimental, and Air Products, which has been abandoned. The time scale for commercialization of these systems is uncertain, but it would be unlikely that any would be commercially available before 1980.

In general, hot gas cleanup processes are not expected to be as flexible as cold liquid scrubbing processes in achieving low-sulfur levels (below 100 ppm) in the product gas(48). This could cause a problem in some industrial situations where very low sulfur levels are necessary to minimize corrosion in gas distribution systems and minimize effects on products in direct-fired furnaces. In this study a sulfur level of 300 ppm was assumed adequate for both environmental, piping, and product degradation purposes. After actual trial or new standards, however, it may be determined that a lower sulfur level would be desired. Under these circumstances, a cold liquid scrubbing system would be more flexible in being able to achieve a lower sulfur level.

None of the hot-gas cleaning systems discussed is capable of removing ammonia and only one, the Battelle-Northwest molten salt, is capable of removing particulates; however, even this process would require filtration of the molten salt, which is a difficult problem yet to be solved. In cold gas liquid scrubbing processes, ammonia and particulates are reduced to low levels in the gas by the water scrubbing steps preceding desulfurization.

The ammonia compounds, if left in the gas, could lead to unacceptably high NO_x emissions for some gasification processes due to oxidation of fuel bound nitrogen (see Table 33 in Section VI). Also, ammonia compounds could lead to higher corrosion rates in piping (see Section V). At present, no processes are available for removing ammonia compounds along with sulfur from hot fuel gas.

Also, a hot fuel gas would result in a higher flame temperature than would a cold fuel gas which would increase the production of thermally produced NO_x . Figure 41 shows the effect of fuel temperature on flame temperature. Flame temperature could be reduced by dilution with excess combustion air; however, this would reduce thermal efficiency by increasing stack losses—defeating the purpose of a hot fuel gas.

Data on particulate loading in raw fuel gas are very limited, but, depending on the process, particulate content can be high. Fixed-bed gasifiers would tend to be lowest due to their large coal size and low flow velocities. The Winkler fluidized-bed gasifier reportedly carries from 50 to 75 percent of the ash in the coal over with the raw gas(51). The

Koppers-Totzek entrained slagging process results in about 50 percent of the ash in the coal being carried over with the raw gas with the remaining dropping out as slag. For one case, Koppers indicates particulate loading in the raw gas of 26 g/Nm^3 (11.57 grain/scf)(47).

Particulate removal devices capable of operating on hot-fuel gas at temperatures similar to those of hot-gas desulfurization systems are not well developed. Electrostatic precipitators have been used successfully at temperatures of 255 to 590 C (500 to 1000 F) in the utility industry for controlling fly-ash emissions. Laboratory tests have been conducted on hot precipitators with gas temperatures up to 815 C (1500 F) with removal efficiencies of 90 to 98 percent; however, long-term continuous operation was not demonstrated(52). A novel granular bed filter has been developed with removal efficiencies of greater than 90 percent on particles down to 2 micrometers(50). Other processes such as cyclones and ceramic filters have also been developed for removing particulates from high temperature gases. Plugging and fouling from tar compounds could be a problem in all high temperature particulate removal systems when operating on raw fuel gas from coal. High temperature corrosion from acid gases such as H_2S is also a potential problem.

In cold gas liquid scrubbing desulfurization systems, particulates are removed in the water scrubbing steps preceding desulfurization. These liquid scrubbing systems can be highly efficient in removing particulates to very low levels in the gas stream. Koppers reports that, for an inlet grain loading of 26 g/Nm^3 (11.57 grain/scf), an outlet loading of 0.004 g/Nm^3 (0.002 grain/scf) is achieved with a two-stage venturi scrubber(46). This represents a removal efficiency of greater than 99.9 percent. It is doubtful that a high temperature particulate removal device could be as efficient as cold gas scrubbing. In an industrial situation, where few furnaces would have particulate control devices, the lower particulate removal efficiency would be a drawback of hot desulfurization systems.

With cold gas cleaning systems, waste-heat boilers can be used to recover heat in the raw gas by generating steam. This steam could be used in the industrial plant, for driving pumps and turbines in the gasification plant, or for sorbent regeneration in the cold-desulfurization system. Using waste heat in this manner minimizes the differences in thermal efficiency

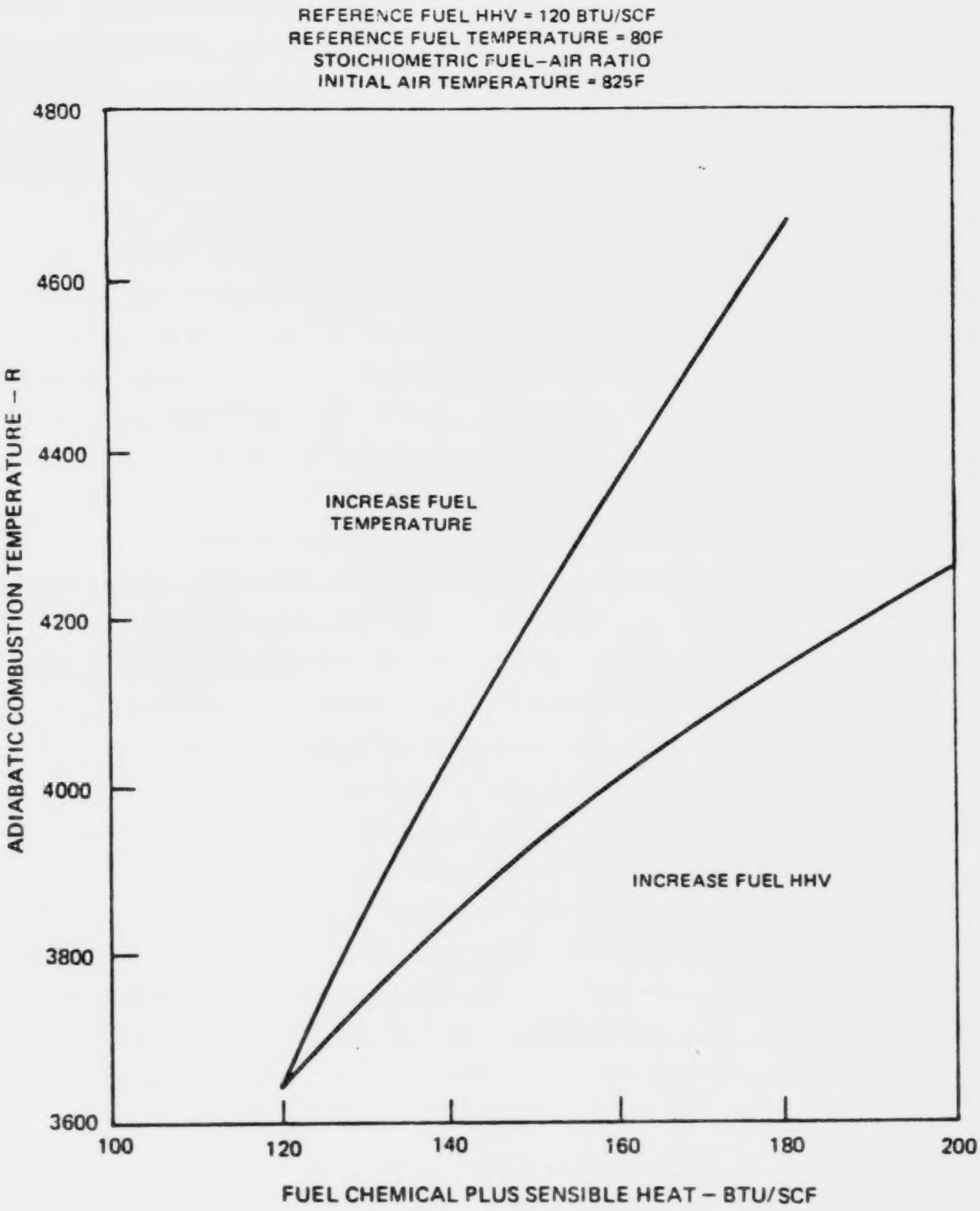

FIGURE 41. EFFECT OF FUEL GAS CHEMICAL AND SENSIBLE HEAT ON COMBUSTION TEMPERATURE(48)

between hot and cold gas processes. Three of the five hot gas desulfurization processes shown in Table 34 (the two dolomite-based processes and the molten salt process) also require steam for sorbent regeneration. The other two, both iron oxide systems, use air and, as a result, yield sulfur as SO_2 which is more difficult than H_2S to convert to a usable or easily handled form. Thus, differences in overall efficiency between hot and cold systems can be minimized with waste heat recovery.

Probably the biggest drawback of hot gas cleaning systems for industrial applications is the necessity for distributing the hot gas in extensive and intricate gas distribution systems often necessary in an industrial plant with a large number of furnaces. As can be seen from Figure 42, a gas temperature of from 705 to 815 C (1300 to 1500 F) would require distribution of three to four times the volume of gas that would be required at 21 C (70 F). In addition, the higher temperatures would increase piping degradation due to corrosion and high stress.

In summary, the availability of a hot gas desulfurization system is not felt to be especially attractive in the industrial situations reviewed here. The inability to remove ammonia combined with higher flame temperatures would result in increased emissions of NO_x. Problems with high temperature particulate removal would also result in increased pollution potential of hot systems over cold systems and, combined with an inability to achieve very low sulfur levels in the product gas, may make hot systems inappropriate in some industrial situations. Also, distribution of a hot gas would magnify corrosion and stress problems in piping and would require larger diameter pipes with the addition of insulation. With waste-heat recovery the difference in efficiency between hot and cold gas systems is reduced (less than 5 percent overall difference in most cases), which would minimize the potential advantage of a hot gas system.

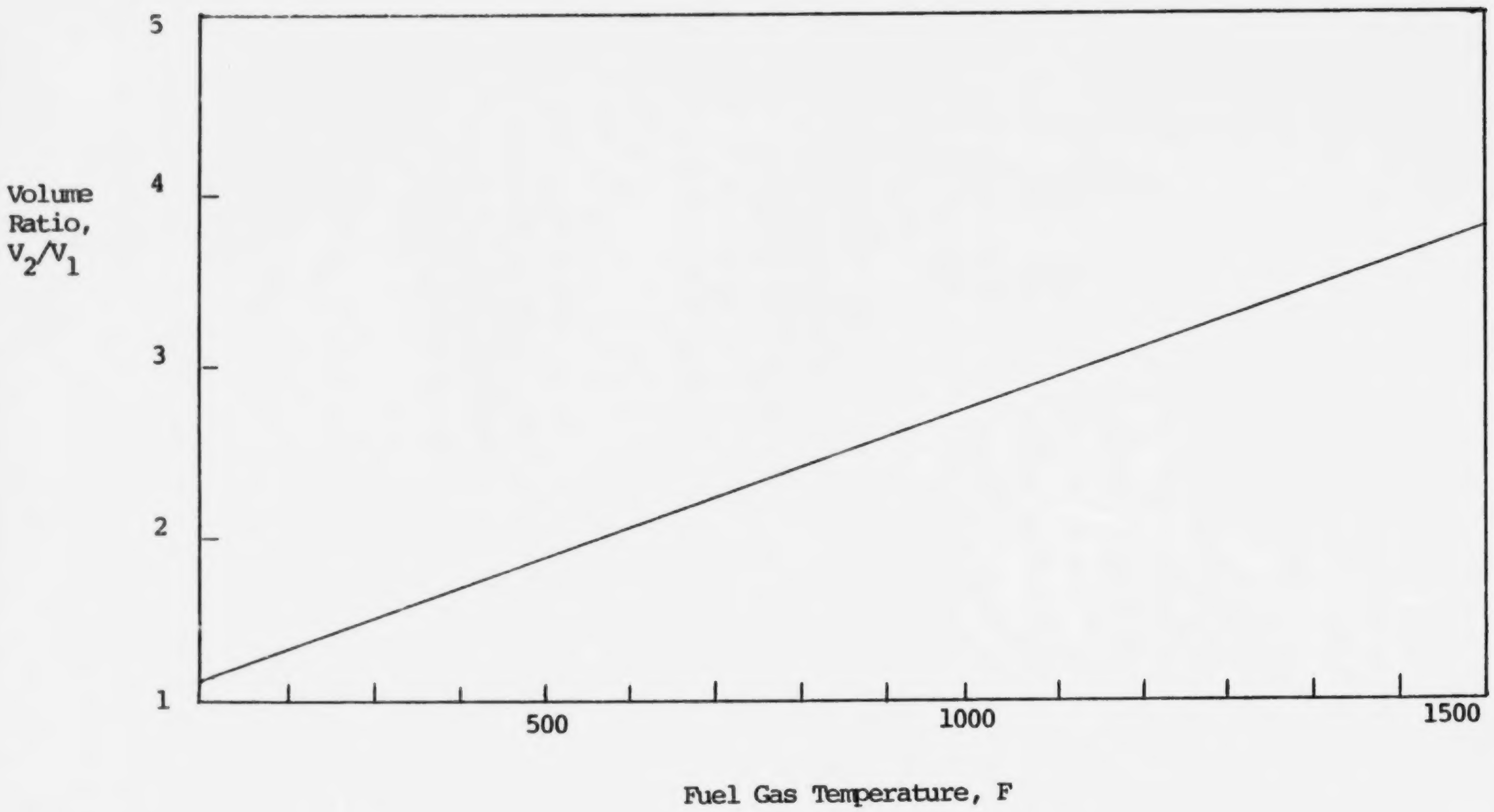

FIGURE 42. RELATIVE VOLUME OF FUEL GAS REQUIRED AT DIFFERENT FUEL GAS TEMPERATURES FOR V_1=70 F

VIII. THE EFFECT OF AVAILABILITY OF ALTERNATE CLEAN FUELS FROM COAL ON INDUSTRIAL DEMAND FOR LOW- AND INTERMEDIATE-ENERGY GAS

A variety of advanced processes currently are under development for manufacturing clean synthetic fuels from coal. These processes are generally more sophisticated than existing commercial units and are intended to operate more efficiently, economically, and on a larger scale. In addition, many of these processes are capable of producing higher grade fuels than low- or intermediate-energy gas.

Table 35 lists those processes currently under development by ERDA for manufacturing substitute natural gas (SNG) from coal. SNG, which has a heating value of about 39.4 MJ/Nm^3 (1000 Btu/scf), is made by methanating synthesis or intermediate-energy gas by reacting CO and H_2 over a nickel catalyst to yield CH_4. The source of the intermediate-energy gas can be most any oxygen-blown gasification process, but the advanced processes under development and shown in Table 35 are intended to maximize the yield of methane in the gasifier to minimize the amount of methanation required. None of the processes shown in Table 35 are expected to be of commercial scale before 1980. A variety of first generation commercial SNG plants are being planned, however, based on current technology. Table 36 lists SNG projects that are currently in advanced stages of planning or awaiting government approvals.

Substitute natural gas from coal is expected to have properties very similar to natural gas and from a combustion standpoint be directly substitutable for natural gas with only minor burner adjustments. Table 37 shows compositions of several natural gases compared to several reported SNG compositions from coal, one from oil, and a sample LNG (liquefied natural gas).

Replacement of Natural Gas by Liquified Natural Gas or Synthetic Natural Gas

It is seen that the normal range of natural gases (even a wider range could be found) brackets the three synthetic gases produced from coal in

TABLE 35. HIGH-BTU GASIFICATION PROGRAM(53)

Major Projects	Contract Value $M (Cost Share)	Contractor	Location	Key Events
• CO_2 Acceptor Process	26.8 (6.6)	Conoco Coal Dev. Co.	Rapid City, S.D.	Methanation plant construction, complete FY 75
• Hygas Process	18.5 (2.0)	Institute of Gas Technology	Chicago, Ill.	Steam oxygen system construction, complete FY 75
Steam-Iron Process	18.2 (7.9)	Institute of Gas Technology	Chicago, Ill.	Complete pilot plant construction FY 75
• Ash-Agglomerating Process	8.9 (1.7)	Battelle-Columbus	West Jefferson, Ohio	Complete pilot plant construction FY 75
• Bi Gas	66.0 (10.0)	Bituminous Coal Research	Homer City, Pa.	Complete pilot plant construction FY 75
• Synthane	19.0	Rust Engineering/ Lumus Corp.	Perc Bruceton, Pa.	Complete construction FY 75

TABLE 36. SNG PLANTS IN ADVANCED STAGES OF PLANNING(54)

Developer	Plant Capacity (10^9 Btu/day)	Expected Starting Date
American Natural Gas Company	1000 × 10^9 Btu/10^9 Btu/day 4-250 × 10^9 Btu/day trains	First train-1981, subsequent trains at 4-year intervals
Cities Service Gas Co. and Northern Natural Gas Company	1000 × 10^9 Btu/day 4-250 × 10^9 Btu/day trains	Currently under study
El Paso Natural Gas Company	785 × 10^9 Btu/day	1980 - first 230 × 10^9 Btu/day plant pending FPC approval
Natrual Gas Pipeline Company	1000 × 10^9 Btu/day 4-250 × 10^9 Btu/day trains	First train-1982, subsequent trains at 3-year intervals
Panhandle Eastern Pipeline Company and Peabody Coal	270 × 10^9 Btu/day	1981
Texas Eastern Transmission Corp. (WESCO)	1000 × 10^9 Btu/day 4-250 × 10^9 Btu/day trains	1980 - first 250 × 10^9 Btu/day train pending FPC approval
Texas Gas Transmission Company	250 × 10^9 Btu/day	1983

TABLE 37. COMPOSITION AND PROPERTIES OF SOME NATURAL GASES, LNG, AND SNG

Composition or Property	Natural Gas No. 1	Natural Gas No. 2	Natural Gas No. 3	LNG	SNG (from oil	SNG (COED)	SNG (Lurgi)	SNG (Bigas)
CH_4	94.9	90.2	72.8	86.3	69.5	88.9	95.8	91.8
C_2H_6	3.1	3.7	6.4	9.0	15.0	0.0	0.0	0.0
C_3H_8	0.3	0.6	2.9	3.2	0.4	0.0	0.0	0.0
Other HC	0.1	0.2	0.6	1.3	0.0	0.0	0.0	0.0
CO_2	1.1	0.8	0.2	0.0	0.3	2.9	2.0	1.1
CO	0.0	0.0	0.0	0.0	0.0	0.1	0.1	0.1
N_2	0.5	4.5	17.1	0.2	0.3	1.6	1.4	1.9
H_2	0.0	0.0	0.0	0.0	14.5	6.5	0.7	5.1
HHV, Btu/ft^3	1029	1009	945	1162	1027	921	872	946
S.G.	0.588	0.609	0.695	0.952	0.541	0.558	0.577	0.549
Stoich. A/F	9.70	9.42	8.90	10.89	9.52	8.63	9.15	8.87
Wobbe No.	1342	1281	1133	1440	1394	1233	1280	1277

respect to Wobbe number, and almost brackets them in respect to the heating value. Thus, one would expect that only minor adjustments would be needed on the control system (5 percent change in Wobbe number is usually assumed to be tolerable without adjustment) to switch to one of these synthetic fuels.

In the case of LNG, an adjustment will certainly be required, resulting from the high ethane content in the fuel. However, the stability limits of the flame are not changed significantly.

A high hydrogen synthetic fuel made from oil is also shown. While the Wobbe number is not as high as that of LNG, the high hydrogen content results in about a 40 percent increased value of the flashback velocity gradient. Thus, there is a possibility with such a fuel as this that premixed burners might have to have their burner faces changed. In precision heat treating, glass forming, and similar operations, the change in flame shape may also result in a need for adjustment when switching to a high hydrogen fuel such as indicated here.

Processes are also currently under development for producing clean liquid fuels from coal by processes termed liquefaction. Unlike gasification, which is an old basic technology, liquefaction is a relatively new technology. Liquefaction of coal was accomplished by the Germans in the 1930's and 1940's using gasification in combination with Fischer Tropsch synthesis, which combines CO and H_2 into higher hydrocarbons from about C_2 through C_6. These lightweight liquid fuels are currently being produced in a large gasification/liquefaction plant in Sasolburg, South Africa, using this type of technology. This technology is generally considered too expensive and inefficient for use in the U.S. today, and so ERDA is funding development of processes for directly hydrogenating solid coal producing a heavy liquid fuel similar to a No. 6 residual oil. Table 38 lists the processes currently being developed to accomplish this.

Another possibly attractive liquid fuel from coal is methanol which is made through gasification in a process very similar to that used for producing SNG (using a copper catalyst instead of a nickel catalyst). The technology for producing methanol from coal is available; however, no commercial plans are known at this time.

TABLE 38. COAL LIQUEFACTION(52)

Major Projects	Contract Value $M (Cost Share)	Contractor	Location	Key Events
• Coal-Oil Energy Development (COED)	21.0	FMC	Princeton, N.J.	Pilot operations complete FY 75
• Solvent Refined Coal (SRC)	41.0	PAMCO	Tacoma, Wash.	Pilot operations started mid FY 75
• H-coal	3.0 (2.7)	HRI	Trenton, N.J. Cattelsburg, Ky.	PDU runs FY 75 pilot-plant decision mid FY 75
• Clean Coke	11.5 (2.9)	U.S. Steel	Monroeville, Pa.	PDU complete FY 77 pilot plant decision FY 77
• Synthoil	(1.1)	Foster Wheeler	Perc	RFP for construction June 75

Table 39 lists some pertinent properties of several grades of oil along with liquified coal, methanol, and shale oil for comparison. When using a liquid fuel as a replacement for natural gas, several considerations are necessary.

Replacement of Natural Gas by Liquid Fuels

In industrial heating boilers and other types of heat exchangers, and in many other industrial applications (note number of dual-fuel burners under secondary steel industry discussed earlier), dual-fuel burners are commonly used which allow natural gas and various grades of oil to be burned simultaneously. These typically burn No. 2 and/or No. 6 fuel oil; in the latter case, provision must be made for heating the fuel slightly to be able to pump and atomize it. It is clear that all the fuels listed in Table 6 except methanol have similar heating values. Thus, if their viscosity is in the proper range (by preheating, if necessary), the fuel nozzle should perform properly at design capacity. The low heating value of methanol indicates that a new nozzle would be required to obtain a higher flow rate. The Bureau of Mines' hydrodesulfurized oil (Snythoil) and the shale oil could be treated as No. 6 oil. It would require preheating by a sufficient amount to be pumped. COED oil which is a product of pyrolysis or gasification falls between No. 2 and No. 4 fuel oil in viscosity, and might require no preheating or only mild preheating, depending on other circumstances. Methanol requires no preheating, but its low viscosity may result in insufficient pump lubrication; thus, a new pumping system might be required as well as new nozzles.

In regard to radiation, all the fuels except methanol would be expected to be highly radiant; those derived from coal would probably be more radiant and might require some dirtying of heat exchanger surfaces (say, by adding magnesium oxide to the fuel) to obtain the proper radiation/convection balance. In the case of boilers, some change might be necessary in superheater controls. Methanol would perform similarly to a somewhat cooled natural gas flame, with low radiant input.

TABLE 39. PROPERTIES OF VARIOUS LIQUID FUELS

Fuel	No. 1 Fuel Oil	No. 2 Fuel Oil	No. 6 Fuel Oil	COED	Bureau of Mines[(a)]	Shale Oil	Methanol
HHV, Btu/lb	19,600	19,400	18,300- 18,700	19,000	17,700	18,000	9,776
Kinematic viscosity, mm^2, at 100 F	1.4-2.2	2.0-3.6		5.5			0.6
at 122 F			92-638		300-500	30	

(a) Bureau of Mines' hydrodesulfurized oil (Synthoil).

In furnaces where dual-fuel burners are not in use, the installation either of such units or of separate liquid fuel burners could be in order. The considerations would be much the same as outlined above, except that the fuel heating system, pumps, and burner could be designed specifically for the fuel. Since, in these instances, it would not be known, a priori, that the change in radiation would have no detrimental effect, this factor would have to be verified.

In some instances, the flame shape is important, and care would have to be taken to ensure a similar shape. Difficulties could be expected with tunnel burners or radiant tube burners; No. 6 fuel oil and similar fuels would not be acceptable in these instances on the basis of presently available burners. Premix burners cannot be replaced directly by fuel oil burners, and alternative burner designs and furnace configurations might be required.

The third alternative is to use a prevaporizer. Liquid fuel is burned on the combustion side of a heat exchanger to heat the main air stream to, say, 370 C (700 F). Liquid fuel is then sprayed into the hot gas and vaporized, and the mixture can be substituted for premixed natural gas and air with minor changes. Systems are commercially available for vaporizing No. 1 and No. 2 fuel oils. It should be noted that methanol will reach a stoichiometric mixture at only 18 C (67 F) mixture temperature, and the fuel rich limit at 40 C (105 F). Air at about 205 C (400 F) would vaporize the methanol to a stoichiometric mixture.

Liquid fuels for the industrial uses studied can be used to completely replace natural gas, provided that at least some fuel with vaporization properties similar to or better than No. 2 fuel oil is available for situations where a gaseous fuel is essential. Also liquid fuels can be stored indefinitely and used when needed, which is an attractive characteristic in industries where fuel demand varies widely from day to day. In all instances but those in which dual-fuel burners are now used successfully, checks would have to be made on the radiation properties. For methanol, because of low viscosity, it might be necessary to change pumps and burner nozzles.

The gasified and liquefied products discussed in this section as alternatives to low- and intermediate-energy gas have attractive features for industry. SNG could essentially be substituted directly for natural gas in practically all industries with almost no foreseeable modifications. Most industries would be willing to pay a somewhat higher price for this fuel over low- or intermediate-energy gas depending on the degree of modification that would be necessary in processing operations. Liquefied coal (No. 6 oil) or methanol could also easily be used in many industries although modifications such as addition or replacement of burners along with installation of heated lines would be necessary. In many industries, however, equipment has already been installed for using No. 6 oil and some operating experience has been gained with its use. In these cases use of liquefied coal would be attractive.

However, the most important factors in determining the potential for use of alternate clean fuels from coal instead of low- or intermediate-energy gas are supply and cost. If all the SNG plants listed in Table 35 were constructed and operated at 100 percent load factor, they would supply about 2.0 x 10^{12} MJ (1.9 x 10^{15} Btu) per year of gaseous energy. In 1972 industry used about 11 x 10^{12} MJ (10.4 x 10^{15} Btu) of natural gas and an additional 58 x 10^{6} MJ (55 x 10^{9} Btu) of oil in supplying its energy requirement of 24.4 x 10^{12} MJ (23.1 x 10^{15} Btu)(1). The total U.S. demand for natural gas in 1972 was 24.4 x 10^{12} MJ (23.1 x 10^{15} Btu) and for oil 34.8 x 10^{12} MJ (33 x 10^{15} Btu)(2). Construction plans and schedules for SNG plants have consistently fallen behind, and today it seems certain that no commercial SNG plants will be in operation by 1980 and only a few by 1985. A total of 176 sites have been identified capable of supporting a 264 x 10^{6} MJ (250 x 10^{9} Btu) per day SNG plant for 34 years(54). If all 176 sites were developed and operated at full capacity, they would produce about 17 x 10^{12} MJ (16 x 10^{15} Btu) per year of SNG. It would be highly unlikely that more than 5.3 x 10^{12} MJ (5 x 10^{15} Btu) per year of SNG could be produced by the year 2000 which would be less than half the 1972 industry demand and less than the 7.7 x 10^{12} MJ (7.3 x 10^{15} Btu) used by the household and commercial sector in 1972(56).

It would be a good assumption, therefore, that SNG or any natural gas equivalent will be reserved for high priority uses in the future and will not be of any higher availability to industry than natural gas currently is.

At present, no commercial coal liquefaction plants have been planned, and only one significant sized demonstration plant is scheduled for construction (to be built in New Athens, Illinois, by Coalcon) producing 6.2×10^5 liters/day (3900 barrels/day) of oil (or approximately 24.3×10^6 MJ per day [23×10^9 Btu/day]) and about 23.2×10^6 MJ/day (22×10^9 Btu/day) of SNG. This plant is not scheduled for operation until the early 1980's.

ERDA's current plans call for the production of about 9.5×10^{12} MJ/year (9×10^{15} Btu/year) of SNG and liquefied coal products by the year 2000(2). This would be equivalent to about 16 percent of our 1972 use of natural gas and oil of 59.2×10^{12} MJ (56.1×10^{15} Btu).

Even if the goal of 9.5×10^{12} MJ/year (9×10^{15} Btu/year) of these fuels is attained, which would require a significantly accelerated pace over that of today, it would be unlikely that these fuels would be available to industry despite their relative ease of application. Rather, they would be reserved for high priority uses such as home heating, transportation, and chemical feedstocks.

Also, significant engineering and development efforts will be required to perfect the processes for manufacturing higher grade alternate fuels from coal. This, combined with the greater complexity required in processing operations, including the necessity for high pressure operation, in many cases will probably result in significantly higher production costs for these fuels compared to those of low- and intermediate-energy gas made with existing processes.

Therefore, industry's pursuit of low- or intermediate-energy gas, which is generated on site for their needs, seems entirely reasonable as a means of securing both near- and long-range supplies of needed fuel.

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APPENDIX A

COMBUSTION OF LOW- AND INTERMEDIATE-ENERGY GAS IN INDUSTRIAL PROCESSES

INTRODUCTION

Moderate- and low-energy gas obtained from various gasification processes have been suggested as substitutes for natural gas in many industries, including the two that are the subject of this report, namely, the secondary steel industry and the refinery industry. Problems to be considered in making such a substitution are flame stability, fuel cleanliness and pollution tendencies, flame heat transfer characteristics, and overall flow rate (fan capacity). Three of these items are discussed below; pollution problems are covered in Section VI.

Flame Stability

It should be noted that a change to moderate or low-energy gases is the reverse direction to that made decades ago; as natural gas became available, the use of various combustible mixtures from coal-gasification processes were phased out. A similar more recent process occurred in England with the development of the North Sea gas supplies. Generally speaking, these fuel changes were accompanied by changes in types of burners. For instance, in the residential area, the quiet, soft diffusion flame burners in heating units were replaced by the noisier, harder, but more compact premixed flame burners. Unfortunately, such changes have reinforced a connotation that moderate- and low-energy gas implies large combustor systems. Yet, the real reason is that the low burning velocity of natural gas permits the use of premixed burners that lead to more compact designs in the case of household heating applications. This example clearly shows that each potential conversion must be analyzed in detail in order to draw valid conclusions.

Basically, flames may be either of the premixed flame type, wherein the fuel and air are uniformly mixed before entering the combustion zone, or of the diffusion flame type, wherein the fuel and air are separated until they reach the combustion zone. In the latter case, however, the leading edge of the flame surface is premixed locally; in fact, in many recent designs of burners, a small premixed region is purposely formed. Thus, the flame stability is related to premixed flame characteristics. The flames may be either laminar, wherein the rate of mixing (of mass, momentum, and energy) is controlled by the molecular kinetic properties, or turbulent, wherein the rate of mixing is controlled to a significant degree by the turbulence characteristic of the flowing gases in the precombustion region. Most industrial burners have turbulent flames; however, in considering the fine details of flame stability, the laminar flame characteristics usually control the local phenomena.

Practical burners can combine features of both types of combustion. For instance, many premixed burners use fuel-rich mixtures; secondary air is added to the products of the premix flame to produce a diffusion flame. Nozzle-mix burners (for example, a fuel jet surrounded by multiple air jets firing into a burner tile) may show either a diffusion flame or premixed flame, depending on where the flame is stabilized. Thus, it is difficult to single out one feature of a combustible mixture that can be considered to characterize the fuel for comparison purposes, even if the burner is not changed in the process of changing fuels.

If a comparative parameter must be chosen, however, the most easily available pertinent parameter seems to be the flash-back velocity gradient. Experimental values of this parameter are obtained by firing a Bunsen-type burner in the open. The flow rate of the combustible mixture to a laminar flame is slowly reduced until the flame flashes back. It is found that the velocity gradient at flashback in laminar flow is independent of duct sizes over a wide range of sizes and ambient atmospheres. Values are available from one source (A-1) of information for a wide variety of fuels, and some combination rules have been developed for those fuel mixtures not listed (for instance, see Reference A-2).

The great significance of the flash-back velocity gradient in studies of industrial combustors is that it is related closely to several other significant combustion parameters. For instance, the flash-back velocity gradient is proportional to the blow-off limit in an enclosed system, to the chemically controlled reaction rate per unit volume, to the square of the burning velocity, and it is inversely proportional to the ignition delay time mentioned by many investigators. It also has been suggested that it is proportional to the peak frequency of the combustion roar spectrum(A-3).

Presentation of Flame-Stability Data

Table A-1 presents the information on the composition and higher heating values of compositions that are considered characteristic of the various fuels considered in this analysis. Table A-2 presents computed values of the stability limits considered from three points of view:

(1) The usual critical flash-back velocity gradients at stoichiometric and the maximum flash-back velocity gradients are presented.

(2) The stoichiometric and maximum values of the heat release rate (the product of the gradient and the corresponding higher heating value per cubic foot of fresh mixtures) are given.

(3) The stoichiometric and maximum value of the products of the gradients and the corresponding higher heating value of the fuel are tabulated.

Also included are the Wobbe Numbers (the high heating value divided by the square root of the specific gravity) which comprise a useful parameter in evaluating fuel changes in aspirating-type premix burners or burners in which pressure sensitive controls are used to regulate the relative rates of flow of fuel and air.

TABLE A-1. FUEL COMPOSITION AND THERMAL PROPERTIES

Gasifier	Gasifying Medium	Volume Percent						HHV, Fuel		Heat Release for Stoichiometric Mixture		Adiabatic(a) Flame Temperature	
		N	CO_2	CO	H_2	CH_4	C_3H_8	MJ/Nm^3	(Btu/scf)	MJ/Nm^3	(Btu/scf)	K	F
Lurgi	Oxygen-steam	1	33	13	37	16	0	12.7	(322)	3.4	(87)	2104	(3328)
Lurgi	Oxygen-steam-stripped	2	0	20	55	23	0	18.7	(474)	3.8	(96)		
Koppers-Totzek	Oxygen-steam	2	7	56	35	0	0	11.6	(294)	3.7	(93)	2320	(3717)
Koppers-Totzek	Oxygen-steam-stripped	1	0	61	38	0	0	12.6	(319)	3.7	(95)		
Coke oven(b)		4.6	0.1	10.6	58.4	26.3	0	19.3	(490)	3.7	(95)		
Wellman-Galusha	Air-steam	50	3	29	15	3	0	6.8	(172)	2.9	(73)	2041	(3214)
Natural gas(c)		0.6	0.9	0	0	91.5	1.3	42.0	(1066)	3.8	(97)	2232	(3358)
Propane(d)		0	0	0	0	0	98.6	99.3	(2521)	4.0	(102)		

(a) Calculated with dissociation, at stoichiometric ratio.

(b) Bureau Mines RI 5225, Fuel No. 43.

(c) Also contains 5.2 percent C_2H_6, 0.5 percent other hydrocarbons; Bureau Mines RI 5225, Fuel No. 1.

(d) Also contains 1.4 percent C_3H_6; Bureau Mines RI 5225, Fuel No. 3

TABLE A-2. FUEL STABILITY FACTORS

Gasifier	Gasification Medium	Flash-Back Velocity Gradient, sec^{-1}		Heat Release Rate, 10^3 MJ/Nm^3-sec (10^3 Btu/ft^3-sec)		Gradient x HHV, 10^4 MJ/Nm^3-sec (10^4 Btu/ft^3-sec)		Wobbe[(a)] No.
		Stoichiometric	Maximum	Stoichiometric	Maximum	Stoichiometric	Maximum	
Lurgi[(c)]	Oxygen-steam	767	775	2.6 (66)	2.6 (67)	1.0 (25)	1.0 (25)	368
Lurgi[(c)]	Oxygen-steam-stripped	1930	1950	7.2 (183)	7.4(187)	3.6 (91)	3.6 (92)	769
Koppers-Totzek[(c)]	Oxygen-steam	2020	2640	7.4 (187)	9.4(40)	2.5 (63)	3.3 (84)	353
Koppers-Totzek[(c)]	Oxygen-steam-stripped	2660	4430	10.0 (254)	15.0(380)	3.4 (86)	5.6(142)	404
Coke oven[(b)]		2200	2290	8.0 (204)	8.5(2.5)	4.1 (105)	4.3(110)	847
Wellman-Galusha[(c)]	Air-steam	584	650	1.7 (43)	1.8 (46)	0.4 (9.7)	0.4 (11)	187
Natural gas[(b)]		420	420	1.6 (40)	1.6 (41)	1.7 (44)	1.7 (44)	1364
Propane[(b)]		560	600	2.1 (53)	2.4 (61)	5.6 (142)	6.0(152)	2019

(a) Higher heating value of the fuel divided by the square root of the fuel specific gravity.

(b) Flash-back velocity gradient obtained from Bureau Mines RI 5225.

(c) Flash-back velocity gradient computed using Reference A-2.

The gradient values of Table A-2 are obtained from Figures A-1, A-2, and A-3. Figure A-1 presents the flash-back velocity gradients as a function of the fuel gas concentration relative to the stoichiometric value.* These gradients were constructed using a modification, presented in Reference A-2, of the techniques presented in Reference A-1 and data from the same source. Figure A-2 presents the critical value of the heating rate per unit volume, based on the fresh mixture properties. Figure A-3 presents the curves of Figure A-1 in an alternate form, each curve being multiplied by the corresponding higher heating value for the fuel. It is noted that the natural gas curve peaks close to stoichiometric, while the remainder of the fuel-air mixtures peak on the fuel-rich side.

A consideration of Figures A-1 and A-3 shows that natural gas behaves much like the fuels that have the lower HHV. Other than natural gas, the produced fuels (principally consisting of H_2, CO, and inerts) line up roughly in order of the amount of inert present. Considering Figure A-1, natural gas (with no appreciable inerts) has stability limits lower than any of the listed fuels resulting from various coal gasification processes.

In some uses of low- and medium-energy gas, the gas may be preheated. Similarly, there are instances wherein the air is preheated. These cases may be analyzed in a manner similar to that discussed below for the nonpreheated cases. However, suitable stability curves similar to those in Figures A-1, A-2, and A-3 must first be constructed.

Discussion of Flame Stability in Burners

Three general types of burners are considered in the discussion of flame stability - premix burners, delayed-mixing burners, and nozzle-mixing burners. While it is not difficult to distinguish premix burners from the other two types, the distinction between delayed-mixing burners and nozzle-mixing burners is sometimes rather vague. For the purpose of this discussion, combustion in a nozzle-mixing burner will be more intense,

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* Relation of fuel gas concentration relative to stoichiometric, F, to air to fuel equivalance ration, α, is given in table of symbols, page A-32.

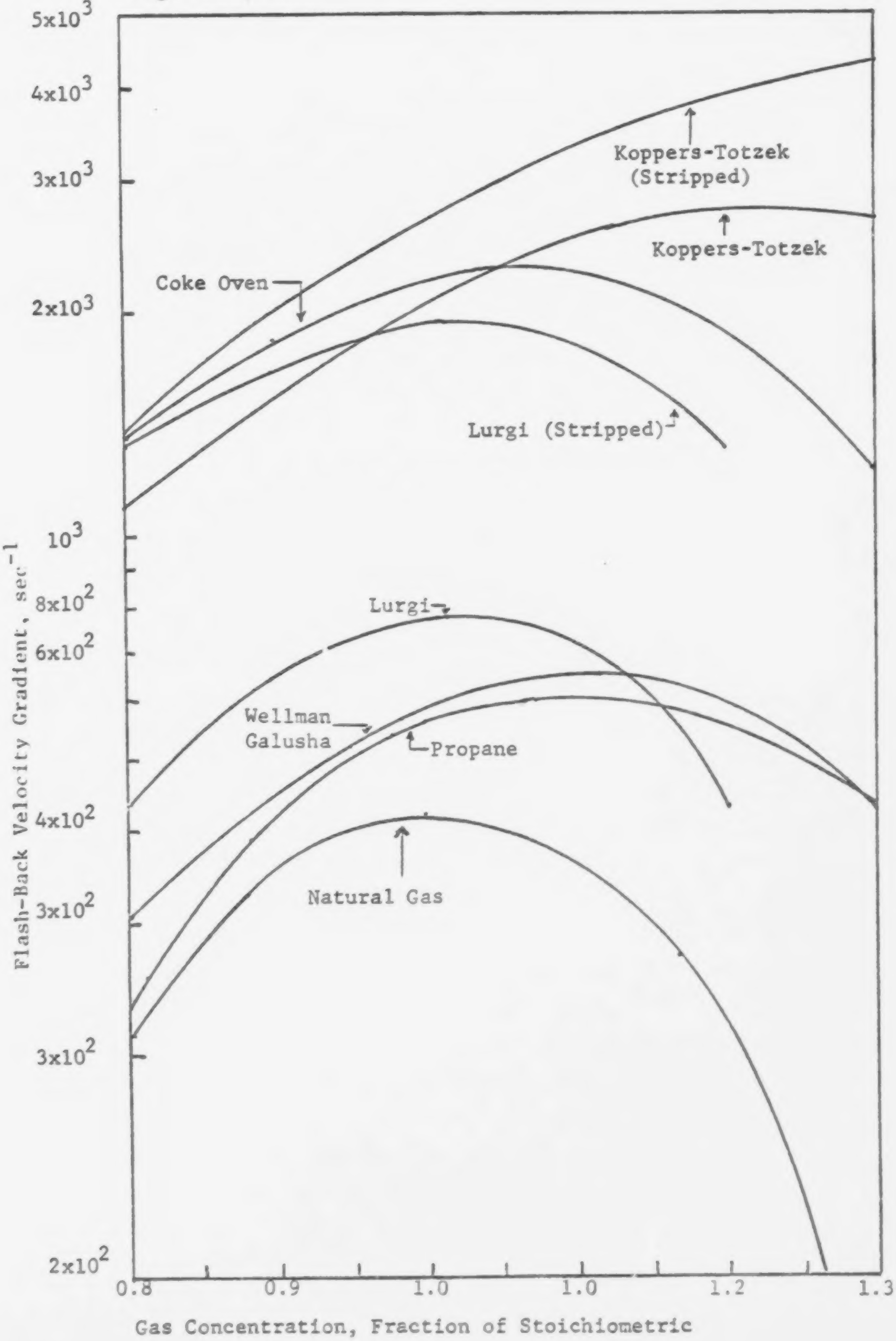

FIGURE A-1. FLASH-BACK VELOCITY GRADIENT AS A FUNCTION OF GAS CONCENTRATION IN MIXTURE

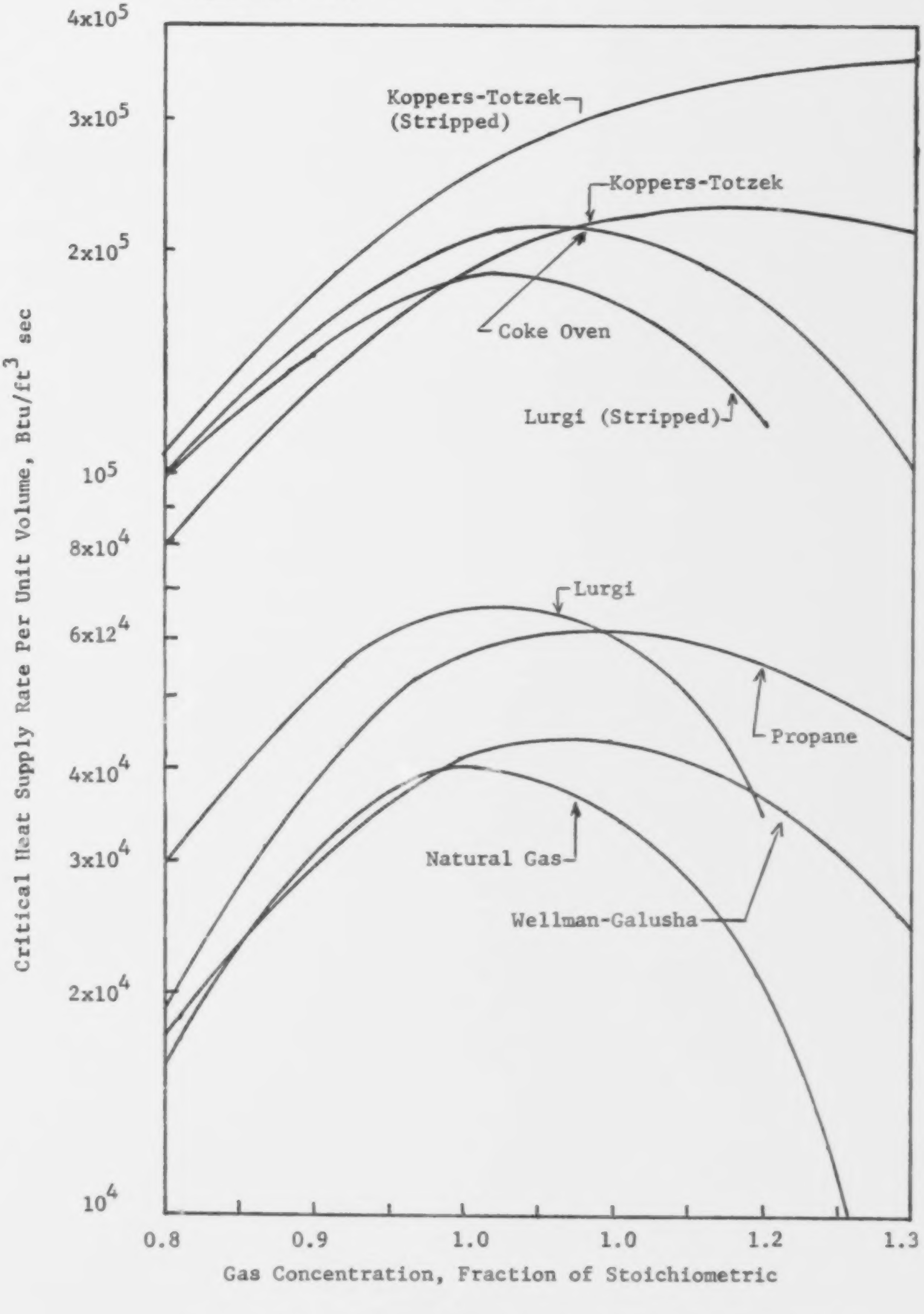

FIGURE A-2. CRITICAL HEAT RELEASE RATE PER UNIT VOLUME (FLASH-BACK VELOCITY GRADIENT TIMES HHV OF MIXTURE) AS A FUNCTION OF GAS CONCENTRATION IN MIXTURE

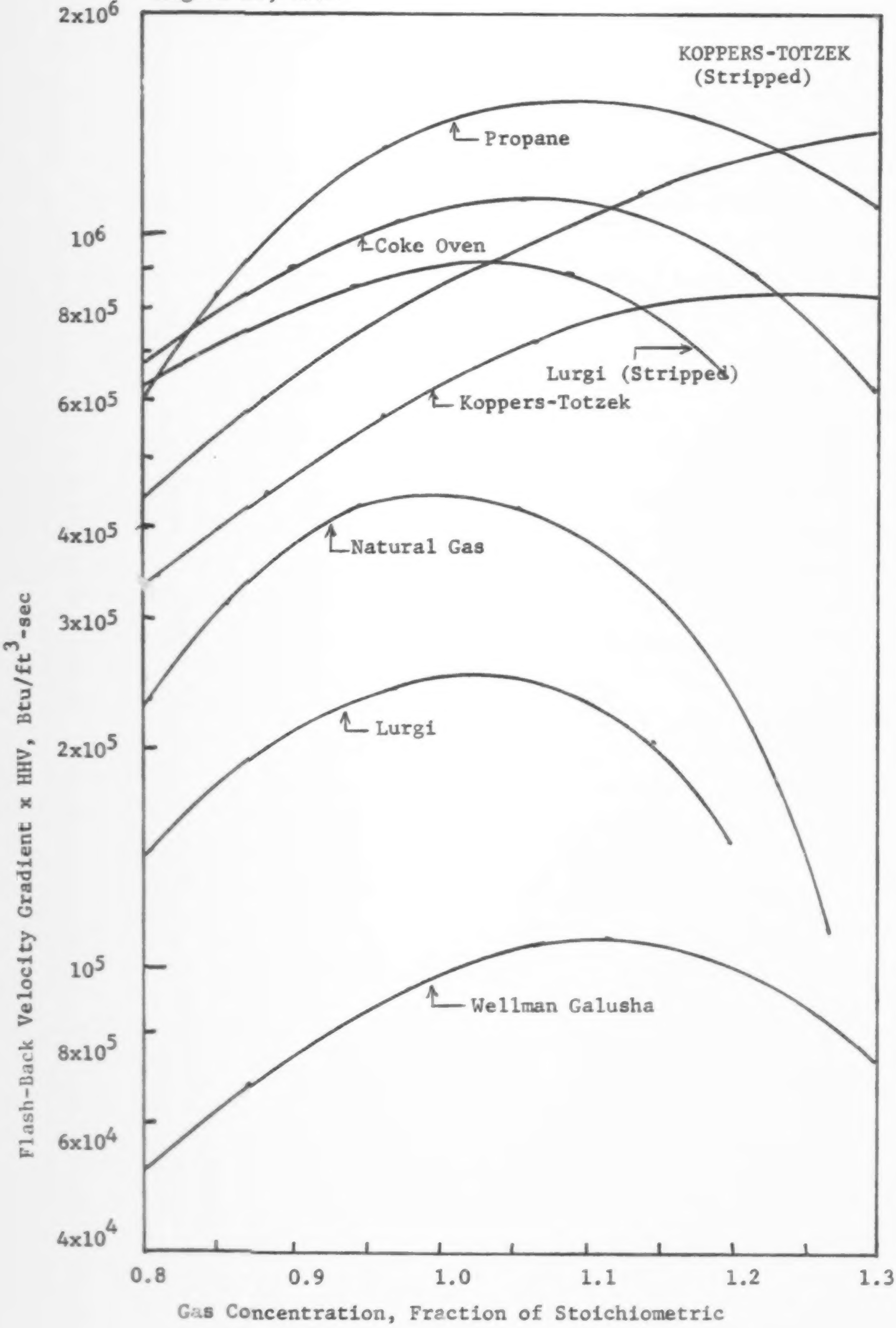

FIGURE A-3. FLASH-BACK VELOCITY GRADIENT TIMES GAS HIGHER HEATING VALUE (HHV) AS A FUNCTION OF GAS CONCENTRATION IN MIXTURE with at least a significant fraction of the combustion taking place within or close to the burner tile. Delayed-mixing flames will generally extend a considerable distance from the burner and often be characterized by a low turbulence level and mixing rate. Significant amounts of furnace gases might be recirculated into the base of their flames. To complicate the problem further, some premixing is often used in delayed mixing and nozzle-mixing burners to aid in producing a stable ignition region for the flame.

Premix Burners.* Premixed flames are reasonably common in industry and are the easiest to analyze. The premixed fuel and air are usually supplied to the region from an inspirator or Venturi mixer, an aspirator or suction-type mixer, or a fan mixer. The burner may be a small port or ported manifold type, a large port (or pressure type), a tunnel burner, or a flame-retention type pressure burner. For high firing rates with turbulent flow, the flame will not hold at the end of the duct. Therefore, a variety of flame-holding systems are used. Figure A-4 is an example of the flame-retention type burner, in which some of the combustible mixture is slowed down and diverted into an annular combustion region. The flame in the protected annular region acts as a pilot to maintain or hold the main flame. In closed systems (such as tunnel burners), steps, recesses, grids, and other obstacles are often used to hold the flame. These form protected recirculation zones, which hold the flame and from which the flame spreads.

In all of these cases, the key factor is a term proportional to the velocity gradient at flashback.** As a simple example, consider a closed

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* "Usually, a burner applied with gas and air which has previously been mixed, but sometimes a burner within which the gas and air are mixed before they reach the nozzle (as opposed to nozzle-mixing burners)."(A-4)

** Often, in the case of flame holding by obstacles, an explanation of performance based on the concept of a delay time is advanced; this delay time is proportional to the reciprocal of the flash-back gradient.

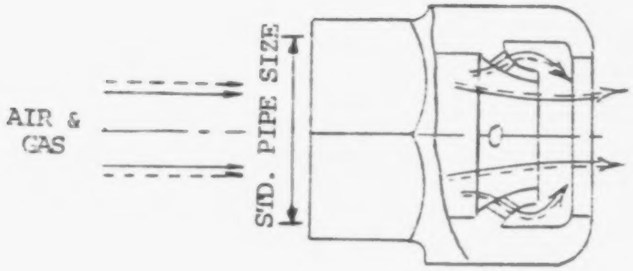
AIR &
GAS
STD. PIPE SIZE

FIGURE A-4. PREMIX BURNER, FLAME RETENTION TYPE system where the blow-off velocity gradient, G_{bo}, is, say B times the flash-back velocity gradient, G_{fg}.[*] For turbulent systems, the gradient is usually given merely in the form of U/D--the average flow velocity, U, divided by a characteristic diameter, D. Then, $U_{bo} = B\ U_{fb} \sim DG_{fb}$. Now, if the critical velocity gradient is doubled by a change in fuel, then both the blow-off and flash-back velocities will double. In many burners, the equivalent of single or multiple steps are used, so that on premature flash back the characteristic diameter is decreased as the flame moves upstream; this decreases the critical flash-back velocity at the same time as the flow velocity increases, thus stopping the flash back. For such a design, increasing the critical value of G will increase the range of flow rates for stable flames, but not necessarily the heating rate, as will be shown.

A constant heating rate system will now be considered with a change in fuel. Considering a single burner with a volume flow rate of combustible mixtures, Q, and a heating value per unit volume of mixture, H_m, the heat release rate will be QH_m. When the critical blow-off condition is reached, the heat release rate is given by $QH_m \approx UD^2H_m \approx D^3GH_m$. For a single size of burner, the key term for comparison is GH_m. The relative values of this term are plotted on Figure A-2. It is seen that, on the basis of the heat release rate at blow-off, natural gas and Wellman-Galusha gas are about the same on the excess air side and all other gases shown here are more stable against blow-off. On the other hand, these gases are more prone to flash back, and their use could result in a significant decrease in turn-down ratio.

If the fan power is limited,[**] the change from, say, natural gas to low energy gas may be complicated by this power limit. The air power is given by $Q\Delta p$, which varies with $\rho Q^3/D^4$. Assuming a constant heat release rate, that is, if QH_m is constant, and that dynamic (rather than viscous) pressure losses are controlling, the air power varies with $\rho/D^4{H_m}^3$. For a stoichiometric mixture in the air, ρ does not vary appreciably in comparison to H_m^3. Thus, the relative values of H_m are of great importance. It is seen from Table A-1 that the values vary from 102 for propane to 73 for gas from Wellman-Galusha gasifier. The stripped Lurgi, the unstripped, and the stripped Koppers-Totzek, and the coke oven gas are directly interchangeable with natural gas on this basis. We note that, if D is increased to compensate for the lower value of H_m of the Wellman-Galusha gas, flashback will be encouraged. (This is the reason that in shifting to a medium- or low-energy gas from natural gas, there is a tendency to shift to nozzle-mixing or delayed-mixing burners).

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[*] A list of symbols used in this section is presented on page A-32.

[**] Similar results are obtained if fan pressure is considered to be controlling.

If the number of burners (or the number of elements in some burner designs) can be changed, then the constant term is $\rho/N^2D^4H_m^3$ (again assuming dynamic pressure losses). Assuming that burner designs for comparative fuels are to be limited by the critical velocity gradient, then the constant term is $\rho G^{4/3}/N^{2/3}H^{5/3}$. It is seen that the number of burners (or number of ports) must be increased in changing to the moderate or low energy gas from natural gas to avoid flashback while at the same time the total area of the burners (ND^2) must vary with $(H_m)^{-3/2}$. Thus, if N is about 1 for natural gas, then N would be about 25 to 30 for moderate energy gases and about 3 to 6 for low-energy gases.

Although the volume of products is not exactly proportional to the volume of fresh mixture, it is close enough that the term $\rho/D^4H_m^3$ can be considered also as a measure of flow power loss through a furnace and stack. Thus, again, while several of the gases, as listed above, are interchangeable with natural gas, Wellman-Galusha gas will require more than twice the pressure compared to natural gas to move the products of combustion through a furnace. In a boiler-type furnace, the higher velocities associated with a change to low heating value gas would result in a higher heat fluxes initially and possibly excessive cooling of the products of combustion in the latter part. In a furnace using direct heat conduction to a material, this could be an advantage.

The Wobbe Number, which is the ratio of the higher heating value of a fuel to the square root of the specific gravity of the fuel, is the common measure of interchangeability in simple combustion units with a fixed firing rate, where (a) fuel is used to aspirate air (inspirator or Venturi-type unit), (b) air is used to aspirate fuel (aspirator or suction type), or (c) a pressure-type control is used to control the ratio of the fuel gas and air. The reasoning that leads to the Wobbe Number is as follows.

Consider a unit in which the fuel is used to aspirate the air. In this combustion unit of fixed configuration, with, say, a constant pressure drop on the fuel spuds, $\rho_f Q_f^2$ is a constant. For the heat release rate to be constant, $Q_f H_f$ is also constant, and it follows that H_f^2/ρ_f is a constant. Normalizing the fuel density to specific gravity and taking the square root results in the Wobbe Number. Therefore, if the Wobbe Number changes, the firing rate of this simple type of unit changes with change in fuel unless spud size or supply pressure is changed.

But this is not the entire story. In a combustion unit of fixed configuration, with any of the types of interconnections between fuel and air mentioned above, the ratio of momentum flux of the fuel to air remains constant. Thus, $\rho_f Q_f^2/\rho_a Q_a^2$ is a constant. If α denotes the air/fuel ratio relative to stoichiometric air/fuel ratio and H_a is the heating value of air, $Q_a H_a = Q_f H_f/\alpha$. By substitution, it follows that $[H_f/(\rho_f/\rho_a)^{1/2}]/(\alpha H_a)$ is a constant. As the heating value of the air that is used in burning any hydrocarbon fuel does not vary greatly, a change in Wobbe Number also results in a change of excess air in the combustor if no other change is made. It is often assumed that a change in Wobbe Number of more than 5 percent requires a change in spud size.

From Table A-2, it is clear that any change from natural gas to one of the other fuels will necessitate a change in spuds or readjustment of the control system in some manner.

Delayed-Mixing Burners[*]. Turbulent mixing is usually considered as the rate controlling factor in turbulent diffusion flames of industrial importance. The chemically limited reaction rate, which is far greater than the gross reaction rate of the furnace, is not considered to be controlling or even important, other than through its effect on flame stability. However, the effect of turbulence itself is not well understood in complex flow systems, and additional complications arise from the presence of a flame that adds a random set of volume sources as the gases expand by heat from random pockets of combustion.

Nonturbulent and turbulent diffusion flames have one feature in common: the flames must be held at some point, line, or area. In a nonturbulent flame, the adjacent fuel and air interdiffuse over the end of the partition separating the two gases. At some distance downstream of the partition, a combustible mixture of varying composition is reached over a region greater than the laminar flame thickness. In this region, at a distance equal to or greater than the quenching distance, a premixed flame develops and holds (or "seats") the diffusion flame. In fact, the diffusion flame may be pictured as a stepwise series of premixed flames, each with hotter but more dilute initial composition.

In a turbulent flame, a firm seating of the flame often does not occur (unless provision for a little local premixing has been properly built into the burner). One notes that local cells of the fuel and air are of different compositions, temperature, and velocities and have different molecular and thermal dilutions as they approach the reaction zone.[**] Thus, there are only local regions where the maximum turbulent flame speed can exceed the velocity of the oncoming fuel-air mixture. Therefore, the flame-holding points shift about in space as the local low-velocity regions shift about in the turbulent stream. Furthermore, all of the leading edges of the flame must move at close to the maximum premixed flame speed through the turbulent mixture, stretching and spreading the flame.* When the flame no longer contains enough local regions where it can "buck" the oncoming stream and not be extinguished, it will blow off unless held by some independent energy source.

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[*] Delayed-mixing burners are those "in which the fuel and air leave the burner nozzle unmixed and thereafter mix relatively slowly, largely through diffusion. This results in a long luminous flame called a diffusion flame, luminous flame, or long flame." (A-4)

[**] This variation from the average of local time and space concentrations is known as the unmixedness of the fluid.

It thus appears that the critical stability parameter in an enclosed turbulent diffusion flame will be related to the maximum flashback velocity gradient rather than the velocity gradient specific to the average mixture ratio.

Figure A-5 shows typical delayed mixing burners that will result in a long luminous flame. Figure A-5a is a version in which the fuel and air velocities are similar and the flow streams are paralled. Increase of the cross-flow gas at the Venturi throat results in a decrease in flame length and luminosity. Natural gas and low Btu gases are interchangeable in this burner with change in gas pressure. We note that a pilot flame is incorporated for ignition and/or piloting of the diffusion flame. The pilot flames are usually premix or nozzle-mix flames. Therefore, if the stability of the diffusion flame depends on the pilot flame, then the stability conditions of the pilot flame are of prime importance. However, even with a pilot flame, the diffusion flame may not be sufficiently held so that a satisfactory flame results. Thus, the stability characteristic of the diffusion flame must also be considered. On the other hand, the pilot flame is not normally subject to a necessity for a turn-down capacity. Current practice in design of burner, for safety

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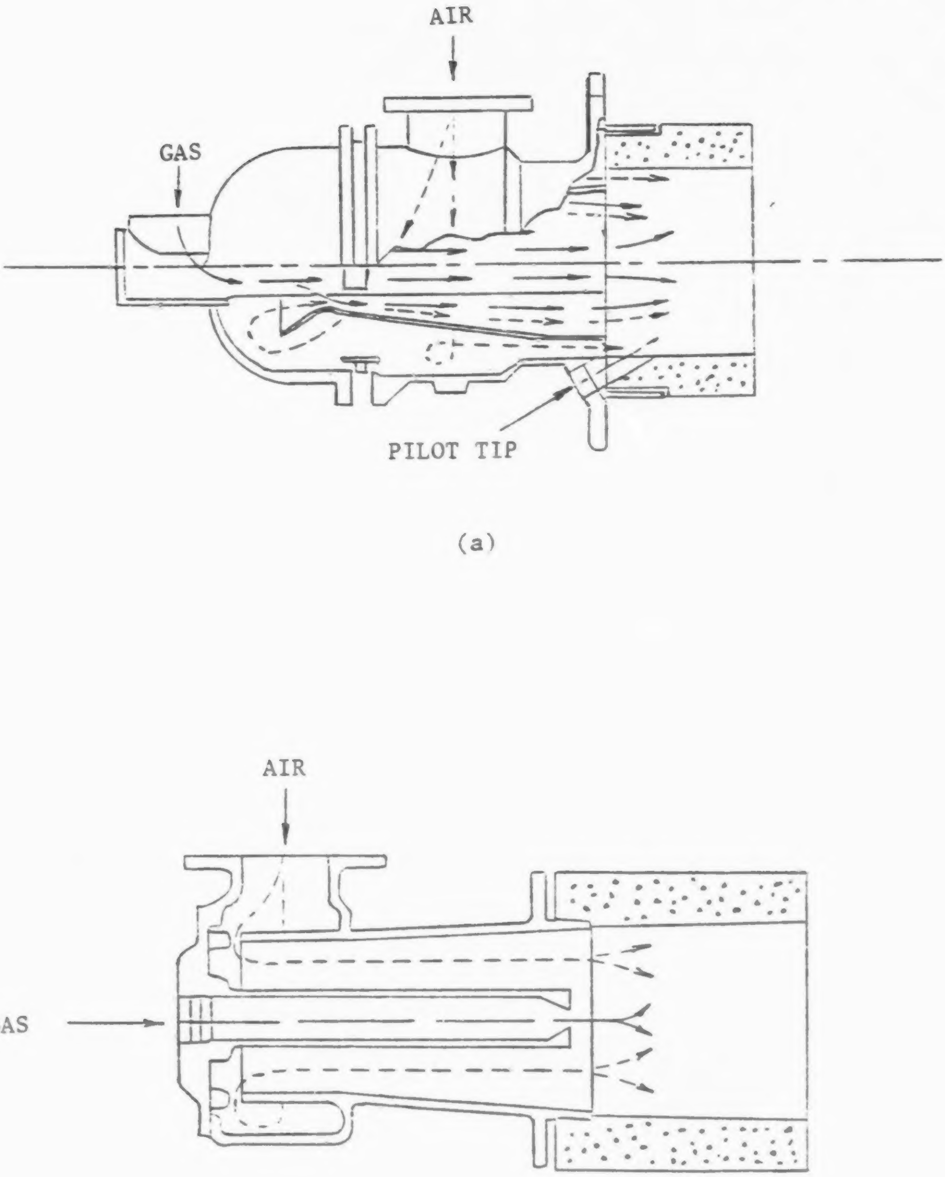
* Otsuka and Niioka(A-5) suggest that, in cases where the flame is being rapidly stretched as would be the case in a turbulent flame front, the flame forms in the maximum temperature region rather than the stoichiometric region often assumed in the literature.


AIR
GAS
PILOT TIP
AIR
GAS

(a)

(b)

FIGURE A-5. DELAYED MIXING BURNERS

reasons, is to insure satisfactory flame performance without a pilot flame. It is noted that the protective effect of the short tile of this burner helps insure satisfactory holding of the flame.

Figure A-5b shows a delayed mixing burner in which the fuel remains in a high-velocity, coherent jet for a considerable distance, surrounded by a low-velocity air mantle. The flame is piloted through the effect of the recirculation and mixing annular region surrounding the fuel jet.

In neither of the burners is there any problem of flashback. Thus, only the possibility of blowing off the flame need be considered in comparing performance with various fuels. Considering the fuels in Table A-1, it is seen from the values for the maximum flash-back velocity gradient in Table A-2 that natural gas is the most unstable of the tabulated fuels. For medium energy fuels, the combustion systems are much more stable. However, this argument does not take into account the necessary change in fuel flow rate with low energy fuels if the burner remains unchanged.

Figure A-5b may be considered as just a simple diffusion flame of the Bunsen burner type, with only fuel in the central jet. With a change in fuel, the maximum diameter of the flame increases as the stoichiometric air/fuel ratio increases. For turbulent flames* at a constant heat input rate, the length of flame changes little. For a constant shape of burner and considering a constant heat input rate and a low velocity of the air in relation to the fuel jet, the holding point of the flame will be determined roughly by the product of the higher heating values and the maximum flash-back velocity gradient. Figure A-3 (and Table A-2) present the values of this point for the various fuels considered. It is seen that the order of fuels has changed from those noted in the previous discussion. Propane, stripped Koppers-Totzek gas and coke-oven gas are the most stable fuels, but natural gas is now above unstripped Lurgi and Wellman-Galusha gases. When the fuel velocity is higher than the average air velocity, the movement of the combustible interface outward with increasing value of the stoichiometric air/fuel ratio also improves the stability of natural gas relative to the remainder of the fuels.

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* While the aspiration rate of the fuel jet cannot be significantly altered, care is usually taken to eliminate as much swirl and turbulence from the air flow as possible to keep from increasing the mixing rate unnecessarily.

This is not the entire story, however. For most delayed mixing burners, such as shown in Figure A-5a, the fuel and air velocities are about the same to inhibit premature mixing.* Therefore, stability of the flame, if the flame is held within the tile, is governed by whichever velocity is controlling--the fuel velocity or the air velocity, or a combination thereof--in the exact region of holding. Furthermore, if changes in fuel are made without concomitant changes in burner dimensions, the relative values of fuel and air velocity will change, and the significant control point may change. The flame may find a stable region or attachment around the annular air jet, rather than the fuel jet. In this case, the flames stabilize closer to the air jet as the air/fuel volume ratio at stoichiometric decreases. Furthermore, and more important, the air velocity does not change much with fuel at a constant heat input rate. In this case, the curves of Figure A-1 should be considered for stability.

If the flame does not stabilize close to the inlets in either position, then the slow mixing can result in other diffusion flames starting beyond the tile in the region where recirculating gases will slow the flow velocity and dilute the air annulus.

When these burners are used in radiant tubes, it is often desirable to have the heat flux peak near the burner and hold at that value or fall off gradually, rather than increase slowly to a peak value some distance down the tube. To accomplish this, a small amount of air may be bled into the fuel jet (or vice versa) so that the boundary of the fuel jet as it emerges from the fuel tube is a combustible mixture. This portion of the boundary burns as a premixed flame, both boosting the heat flux to the wall near the inlet and serving as a pilot for the downstream diffusion flame. However, because of the diffusion effects, there is still a composition gradient, and the stability even in this case should probably be treated as one would a diffusion flame stability problem.

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* These are sometimes called laminar flow burners, but this does not denote viscous flow (Reynold's numbers are still high). Rather, it denotes flow without high intensity turbulence in the interface.

The pressure drops that are involved in supplying the fuel are now considered briefly. As may be deduced from the discussion of the Wobbe Number, in connection with premix flames for a pair of fuels in which this number does not vary too much, the fuels are interchangeable in diffusion-flame applications as well as premix-flame applications. It is seen from Table A-2 that the medium energy gases are the closest to natural gas, but are far from being within the 5 percent variation usually allowed. Furthermore, a massive addition of propane, about 32 percent by volume for the Koppers-Totzek unstrippable gas, would be required to boost the values sufficiently to bring them within range. But it is noted that the energy values of the stoichiometric mixtures are about the same for these fuels as for natural gas, so that changes only in the burner or control settings would be required to obtain satisfactory operation of a burner system.

Interestingly, increasing the orifice sizes for the medium energy gas sufficiently to maintain the same stoichiometry percent results in a decrease in gas pressure while maintaining a constant heat release rate. Changing the orifice size a lesser amount so as to maintain the back pressure on the fuel, and maintaining a constant heat release rate results in an increase in the excess air using fuel aspiration. This, of course, may be handled by an additional adjustment.

One can conclude, therefore, that in replacing natural gas in a diffusion flame with medium energy manufactured gas, no stability problems will be encountered. In confirmation of this, one may note that burner manufacturers often indicate these burners can be used with both natural gas and coke-oven gas. However, there can be a stability problem with lower energy fuels if some changes in burners are not made. For extreme cases, the burner and type of flame may have to be changed.

Nozzle-Mix Burners.[*] Nozzle-mix burners combine the advantage of the relatively short flame of the premix burner and the lack of flashback problems of the diffusion flame. The short flames are obtained by three different methods. Figure A-6a shows the use of multiple high-velocity air jets parallel with the fuel jet. The air jets aspirate the fuel in around them and form short flames because of the small jet diameter and potential core length.[**] Figure A-6b shows the use of nonparallel jets. These may impinge, may interlace (with multiple fuel jets as well as air jets), or may be canted to produce a swirl flame and even a heavy recirculation zone on the burner axis. If a disk is added to the end of the fuel jet in Figure A-5b, a high velocity air flow and a recirculation zone are formed which lead to an intense mixing. The burner in Figure A-5b then becomes a nozzle-mix burner. Some of the fuel in this case may be diverted radially to improve mixing further. In all these cases, the internally recirculating hot gases plus the hot ceramic tile wall provides good flame stability.[***]

If the flame is held as a diffusion flame in a nozzle mix burner, then the flame might either be held around the central fuel jet or the peripheral jets. The argument here is exactly the same as for the delayed mixing burners. The main difference is that, when the flame is not attached close to the inlet of either the fuel or air, rapid mixing may take place before a stable region for the flame to seat is encountered. In this case, the action of the flame is much like a premix burner.

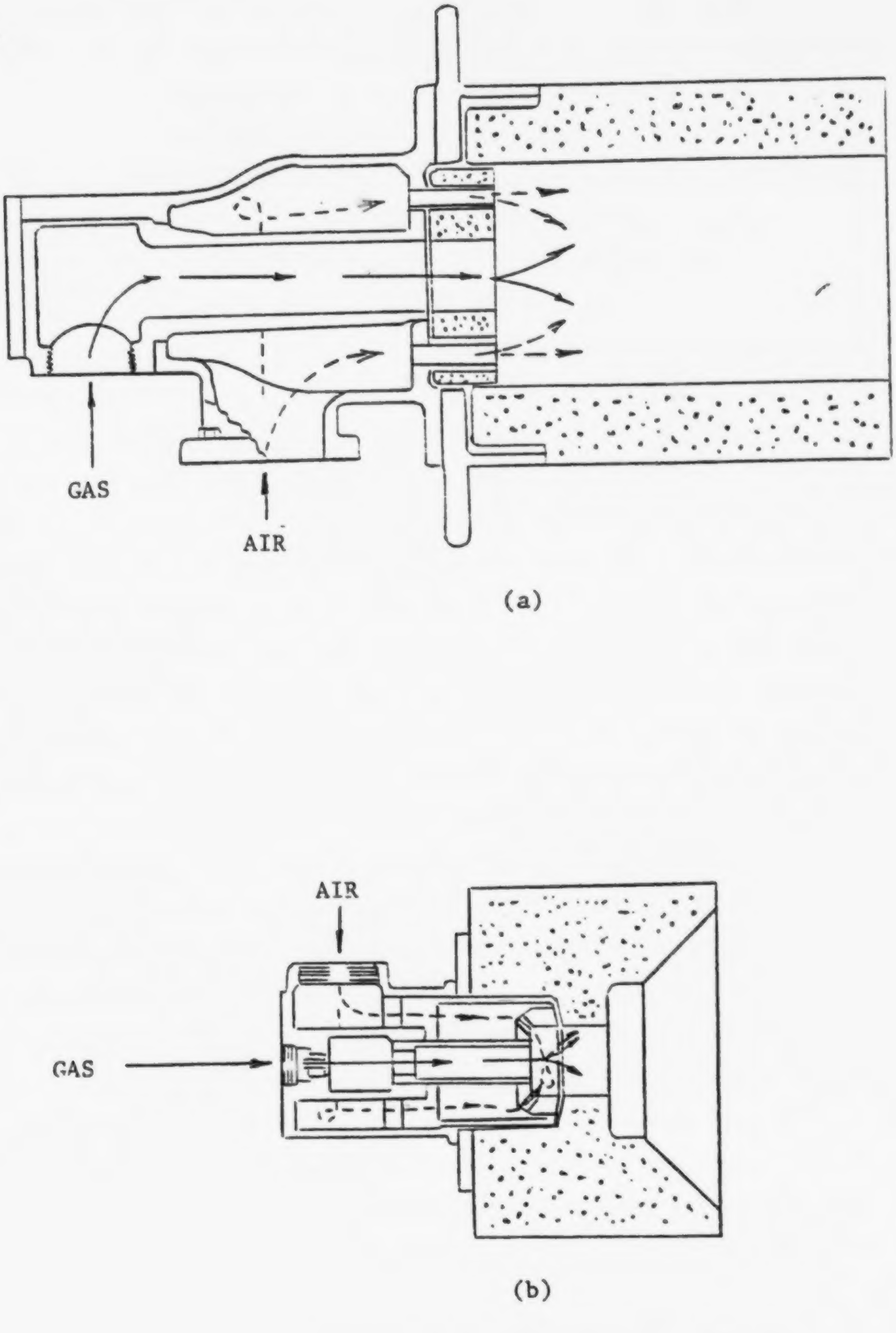
Therefore, it is concluded that in changing from natural gas to moderate or lower energy fuel in a nozzle mixing burner, the position of the flame base may change from around the fuel jet to around air jets, or vice versa, depending on relative flow velocities and change in laminar flame speed. Therefore, an unqualified comparison of stability cannot be made. As a result, it is not clear whether a flame might satisfactorily contain itself within a nozzle-mixing burner tile with a specific change in fuel. Again, as for the delayed-mixing burners, it should be noted that several designs are specified by the manufacturers as operating with either natural gas or coke oven gas.

---

[*] "A burner in which fuel and air are not mixed until just as they leave the burner port, after which mixing is usually very rapid. The flame cannot flash back to this type of burner". A-4.

[**] On occasion, the role of the fuels and air jets are reversed.

[***] Care must be taken to prevent aspiration of cold furnace gases both into the tile and the flame base.


GAS
AIR
AIR
GAS

(a)

(b)

FIGURE A-6. NOZZLE-MIXING BURNERS

Flame Radiation

The effect of change in fuel on radiation output will now be considered. It is obvious that heat is also transferred by convection to work surfaces, and to boiler tubes. As a result, if less heat is transferred by radiation, more heat may be transferred by convection, with a resulting decrease in the overall effect of the change. In furnaces where large amounts of recirculating gas are present, the buffering effect is increased further. Since much of the radiation will come from gases cooled from their maximum temperature, differences in radiation will be reduced by this effect as the gases loose heat. Particulate radiation is ignored in this treatment, first, because there should be a little particulate in the clean gases considered, and second, because no way of estimating an expected concentration is available.

Figure A-7 is generated from Figures 6.9, 6.11, 6.12, and 6.13 of Reference A-6, using product composition and temperature for adiabatic burning with 10 percent excess air of certain of the fuel gases listed in Table A-1 (unstripped). It is interesting to note the high radiating ability of the natural gas flame, for flames more than about one foot thick. Ultimately, of course, all the curves must flatten out at great thickness as they cannot radiate in excess of the black-body temperature of the particular composition. It is also noted that only the K-T gas exceeds the natural gas in radiation, although the Lurgi gas is not too much lower. Stripping of the CO_2 from the Lurgi gas would raise the products temperature and probably bring all three curves close together. The product gases of air-blown gas producers are highly diluted with nitrogen, and as a result, the flame is cooled and the radiation is decreased, as seen from comparing the Wellman-Galusha curve and the Winkler curve with the natural gas curve.

A curve is also shown for the effect of air preheat on the radiation output, for the Winkler gas. It is seen that the radiation output is increased, but far less than enough to bring the gas up to that of natural gas.

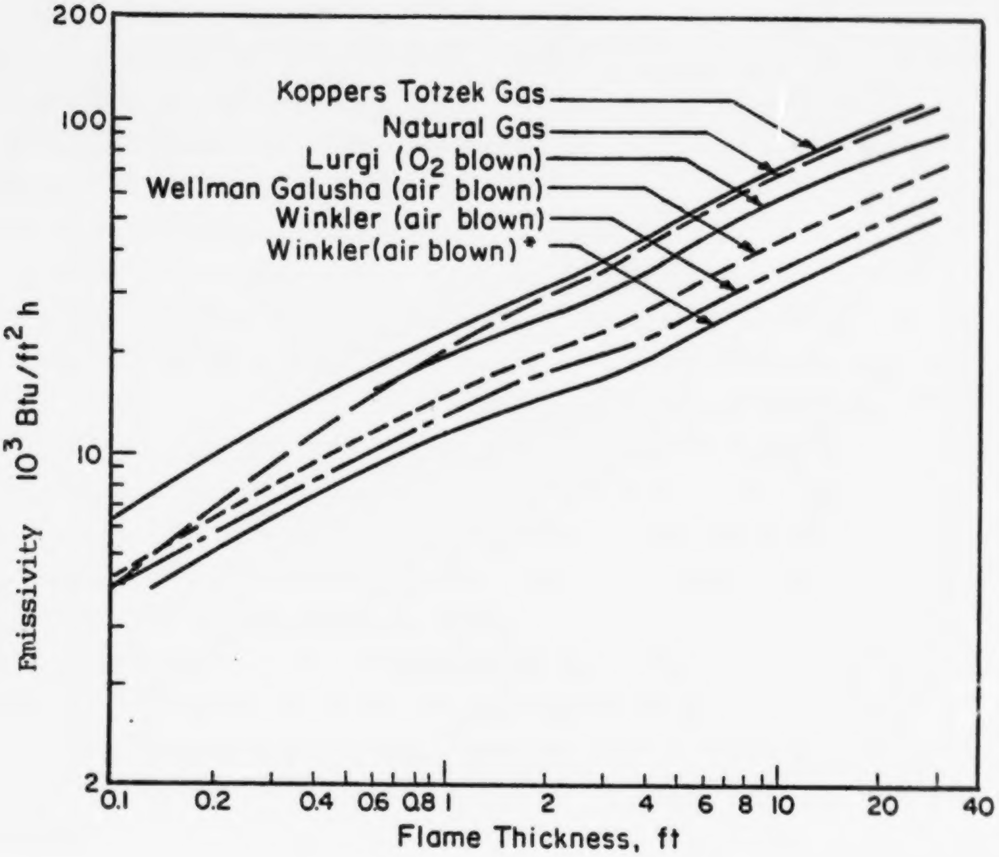


FIGURE A-7. RADIATION FROM ADIABATIC FLAMES AT 10 PERCENT EXCESS AIR

Some feel for the magnitude of the effects resulting from the various changes in the fuel products can be obtained from a consideration of Figures 6-14 of Reference A-6, which is a simplified emissivity chart for CO_2-H_2O mixtures in a restricted temperature range. A temperature-emissivity product is plotted as a narrow band of curves covering a range of ratios of partial pressure of H_2O to CO_2, against the flame thickness times the sum of the CO_2 and H_2O pressures. As an example, radiation from the product gases from stoichiometric combustion of the natural gas and Winkler gas are compared.

The slightly greater amount of (CO_2 + H_2O) for the natural gas leads to about 2 percent greater temperature-emissivity product for natural gas, while the change in H_2O/CO_2 ratio is from 1.90 to 0.41 leads to about 10 percent greater temperature-emissivity product for the natural gas flame (actual amount increases with flame thickness). The absolute temperature ratio of the natural gas to Winkler gas is about 1.14. Thus, even though one temperature term is already in the temperature-emissivity product, there is a further 50 percent increase of natural gas radiation compared to Winkler gas. Thus, the gas temperature itself has the largest effect. As mentioned before, convection heat transfer effects, gas cooling from heat losses, and any soot radiation effects will reduce the significance of these differences, but the differences will still be sufficiently large so that they must be evaluated.

Another aspect of radiation is that associated with flame detection and safety considerations. From the above discussion, it is clear that the performance of any radiation activated controls on a furnace must be considered, if the fuel is changed.

Flow Considerations

There are three different comparisons that might be made relative to flow rate when low or intermediate heating value gas is substituted for natural gas. On the basis of equal heat inputs, the direct substitution of one fuel for another in the fuel lines can be compared. Assuming stoichiometric mixture, the flows of premixed fuel and air can be compared and the product flows can be compared. Table A-3 presents these comparisons, relative to natural gas, for the three replacement gases of immediate interest to the project. Both relative flow velocities, and more important, relative pressure drops (assuming turbulent flow) are given.

If the same fuel lines are used, typical intermediate energy gas from oxygen-blown producers must be delivered to the point of application at 3 to 4 times the flow rate of natural gas to achieve the same heat input. The low energy fuels from air-blown producers require anywhere from 6 to 9 (for Winkler gas, not listed) times the flow of natural gas. The differences in the flow rates of the stoichiometric mixtures are less pronounced than those for the fuel, since the "heating value" of air is about constant. Because of the collapse effect of burning CO or H_2, as compared to hydrocarbons, the product gases may have a lesser volume at standard conditions than the raw mixture. As a result, the product flow rate for K-T gas actually is lower than for natural gas. For the low energy gas from an air-blown producer the increase is less than 20 percent. The corresponding relative increase in pressure drop for the various fuels that would result if the same fuel and flue gas equipment is used is also shown. For intermediate energy fuels about 10 to 15 times the pressure drop would be incurred through existing distribution mains and burners; there is between a negative 20 percent and positive 10 percent change in pressure drop through heat exchangers and other gas passages downstream of the combustion zone. For the low energy fuels, however, pressure drops of over 50 times that for natural gas would be expected in existing mains and burners; corresponding pressure drops in passages downstream of the combustion zone would show a 50 percent or more increase relative to natural gas.

The increased flow rates and pressure drops in fuel supply systems, burners, heat exchangers, and exhaust flues that could be encountered in retrofitting a process from the use of natural gas to low or intermediate energy gas while maintaining the same process heat input could pose a serious problem. Supplying the necessary increased fuel supply rates to various processes throughout an industrial plant will require either pressurized distribution mains, larger distribution mains, or some combination of the two. Pressurized mains would complicate the problem of potential leakage of a toxic carbon monoxide-laden gas into working areas. Increasing the size of distribution systems to handle the increased flow at lower pressures could create problems for processes widely dispersed throughout the plant or in the areas where space is at a premium. Only one of the three gasification systems considered commercial in this study--i.e., the Lurgi process--delivers the fuel gas under pressure (300 to 500 psig, 2070 to 3430/m^2). Fuel gas from the other two processes would have to be compressed, either before or after the gas cleanup stage, for pressurized distribution.

The increased flows and pressure drops occurring downstream of the combustion zone with certain of the substitute gases, though less than those in fuel supply systems, can potentially be a more serious problem. Induced draft and forced draft fans would have to be boosted to higher operating pressures to compensate for higher flow rates. In some cases it may be possible to reduce the pressure drop through the process, such as by removing tubes in boiler heat exchangers, to allow greater volumes of flow at lower pressure drops without upsetting the heat transfer characteristics of the process.

If changes in the process cannot be made to compensate for increased flows and pressure drops, process derating may be necessary. This problem could be more severe for handling the increased volumes of flue gases than for handling of increased volumes of fuel. Analysis of Table A-3 reveals that inability to handle additional flue gas volume could result in a derating of up to 5 percent for intermediate energy (300 Btu/scf; 2664 Kcal/m^3) gas or up to 25 percent for low energy gas (150 Btu/scf; 1332 Kcal/m^3).

TABLE A-3. COMPARISON OF VOLUMES OF FUEL GAS TO NATURAL GAS

Process	Gasifying Medium	HHV, MJ/Nm^3 (Btu/scf)		Relative Flow Rates Fuel Gas/Natural Gas			Relative Pressure Drop Fuel Gas/Natural Gas		
		Fuel Only	Stoichiometric Raw Mixture	Fuel	Stoichiometric Raw Mixture	Stoichiometric Products	Fuel	Stoichiometric Raw Mixture	Stoichiometric Products
Lurgi	Oxygen-steam	12.7 (322)	3.4 (87)	3.31	1.11	1.03	13.5	1.19	1.11
Koppers-Totzek	Oxygen-steam	11.6 (294)	3.7 (93)	3.63	1.04	.885	14.9	1.01	.806
Wellman-Galusha	Air-steam	6.8 (172)	2.9 (94)	6.20	1.31	1.18	52.7	1.66	1.49
Natural Gas		42.0 (1066)	3.8 (97)	1.0	1.0	1.0	1.0	1.0	1.0

Summary

Alterations necessary in burner designs are uncertain without laboratory data on which to base the redesign. Burners that at one time were used for low energy gas are not readily available today as production items. Further, the old designs would no longer be acceptable in most cases, as advances in burner and process technology have resulted in burners with generally wider stability ranges, intermittent instead of continuous piloting, and sophisticated combustion monitoring and control. This aspect of the problem is discussed in a subsequent section.

It is generally felt that adequate industrial burners can be developed for intermediate and most low energy gases. Generally, flow areas in fuel supply lines and burner parts would have to be increased to handle the increased fuel flows necessary to maintain the same energy input as with natural gas or oil. Overall burner diameters or tile diameters probably would not be increased in most cases, however, minimizing the amount of modification necessary in the walls of the furnace.

Larger general-purpose burners should be easiest to retrofit to low or intermediate energy gas. However, the performance of some specialty types of burners may be difficult to duplicate. These include high-intensity or high-velocity burners, burners where particular flame shapes are necessary and applications requiring carefully controlled mixing and combustion rates.

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(A-2) Putnam, A. A., "Effect of Recirculated Products on Burning Velocity and Critical Velocity Gradient", Combustion and Flame, 22, 277-279 (1974).

(A-3) Giammar, R. D., and Putnam, A. A., "Combustion Roar of Turbulent Diffusion Flames", Trans. J. Engineering for Power, 92A, 157-165 (1970).

(A-4) North American Combustion Handbook, The North American Manufacturing Company, 1952

(A-5) Otsuka, Y., and Niioka, T., "The One-Dimensional Diffusion Flame in a Two-Dimensional Counter Flow Burner, Combustion and Flame, 21, 163-176, (1973).

(A-6) Hottel, H. C. , and Sarofim, A. F., Radiative Transfer, McGraw-Hill Book Company, 1967, Figures 6-14.

LIST OF SYMBOLS

D	Characteristic burner diameter
F	Gas concentration, fraction of stoichiometric
G_{fb}	Critical flash back velocity gradient, calculated from data of Reference A-1 by Reference A-Z modification of Reference A-1 techniques
G_{bo}	Blow off velocity gradient
S	Mole fraction of fuel in a stoichiometric mixture
U	Average flow velocity through the burner
U_{fb}	Average flow velocity at flash back, proportion to DG_{fb}
U_{bo}	Average flow velocity at blow off, proportional to DG_{bo}
Q	Volume flow rate of combustible mixture
Q_f	Fuel flow rate
Q_a	Air flow rate
H_m	Heating value per unit volume of combustible mixture
H_f	Heating value per unit volume of fuel
H_a	Heating value per unit volume of air
N	Number of burners
ΔP	Pressure drop across burner
α	Air/fuel ratio relative to stoichiometric air/fuel ratio; $= (1 - FS)/F(1-S)$
ρ	Density of combustible mixture
ρ_f	Fuel density
ρ_a	Air density
B	Ratio of blow-off to flash-back velocity gradient

APPENDIX B

MATERIAL AND ENERGY BALANCES FOR MODEL PLANTS

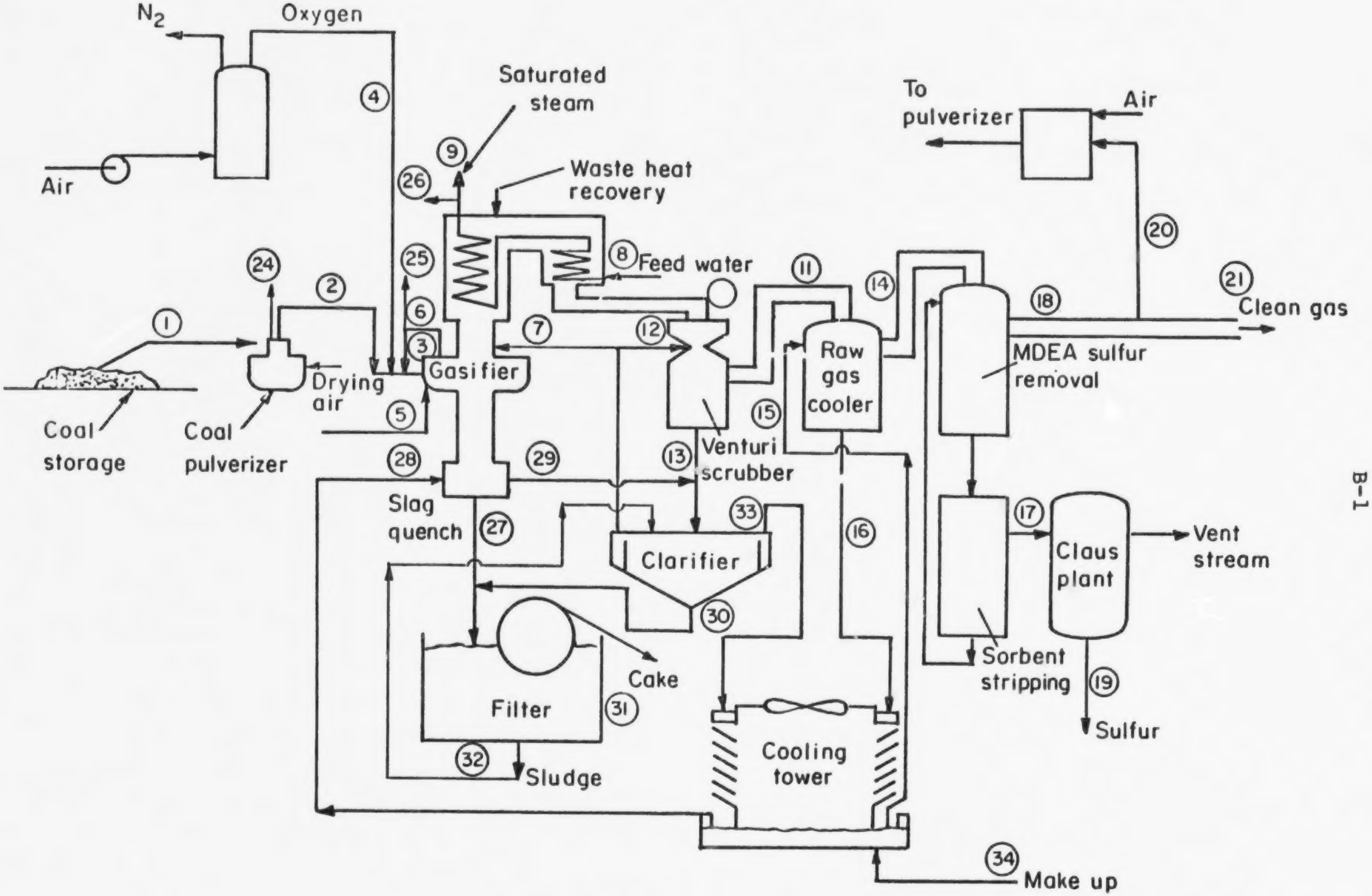
N2
Oxygen
Air
④
Saturated steam
⑨
㉖
Waste heat recovery
To pulverizer
Air
⑳
㉔
②
㉕
⑥
③
⑧
Feed water
⑪
⑭
⑱
㉑
Clean gas
①
Gasifier
⑦
⑫
Raw gas cooler
MDEA sulfur removal
Coal storage
Coal pulverizer
Drying air
⑤
⑮
Venturi scrubber
⑬
㉘
㉙
Slag quench
㉗
㉝
⑯
⑰
Claus plant
Vent stream
Clarifier
㉚
Cake
Filter
㉛
㉜
Sludge
Cooling tower
Sorbent stripping
⑲
Sulfur
㉞
Make up

FIGURE B-1. KOPPERS/MDEA GASIFICATION PLANT MATERIAL BALANCE FOR MODEL STEEL PLANT

TABLE B-1. KOPPERS/MDEA GASIFICATION PLANT MATERIAL BALANCE FOR MODEL STEEL PLANT
(Metric and English Units)

Stream No.	1				2				3			
Description	Coal as Received				Coal to Gasifier				Steam to Gasifier			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	wt%	lb/hr	wt%	Kg/hr	wt%	lb/hr	wt%	Kg/hr	Mol.%	lb/hr	Mol.%
C	34915.7	48.58	76976	48.58	34915.7	63.45	76976	63.45	--	--	--	--
H	2356.9	3.28	5196	3.28	2356.9	4.28	5196	4.28	--	--	--	--
N	352.0	0.49	776	0.49	352.0	0.64	776	0.64	--	--	--	--
S	898.1	1.25	1980	1.25	898.1	1.63	1980	1.63	--	--	--	--
O	7186.7	10.00	15844	10.00	7186.7	13.06	15844	13.06	--	--	--	--
Cl	36.3	0.05	80	0.05	36.3	0.07	80	0.07	--	--	--	--
Ash	7079.7	9.85	15608	9.85	7079.7	12.87	15608	12.87	--	--	--	--
CO	--	--	--	--	--	--	--	--	--	--	--	--
CO_2	--	--	--	--	--	--	--	--	--	--	--	--
H_2	--	--	--	--	--	--	--	--	--	--	--	--
N_2	--	--	--	--	--	--	--	--	--	--	--	--
H_2S	--	--	--	--	--	--	--	--	--	--	--	--
COS	--	--	--	--	--	--	--	--	--	--	--	--
H_2O	19045.4	26.50	41988	26.50	2200.8	4.00	4852	4.00	11807.9	100.00	26032	100.00
O_2	--	--	--	--	--	--	--	--	--	--	--	--
SO_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL												
	Kg/hr	71870.8	lb/hr	158448	Kg/hr	55026.2	lb/hr	121312	Kg/hr	11807.9	lb/hr	26032
	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--
	Kl/hr	--	GPM	--	Kl/hr	--	GPM	--	Kl/hr	--	GPM	--
Temperature	C	--	F	--	C	71	F	160	C	121	F	250
Pressure	atm	--	PSIG	--	atm	--	PSIG	--	atm	2	PSIG	15

TABLE B-1. (Continued)

Stream No.	4				5				6			
Description	Oxygen to Gasifier				BFW to Gasifier Jackets				Steam from Gasifier Jackets			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%
C	--	--	--	--	--	--	--	--	--	--	--	--
H	--	--	--	--	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--	--	--	--	--
S	--	--	--	--	--	--	--	--	--	--	--	--
O	--	--	--	--	--	--	--	--	--	--	--	--
Cl	--	--	--	--	--	--	--	--	--	--	--	--
Ash	--	--	--	--	--	--	--	--	--	--	--	--
CO	--	--	--	--	--	--	--	--	--	--	--	--
CO_2	--	--	--	--	--	--	--	--	--	--	--	--
H_2	--	--	--	--	--	--	--	--	--	--	--	--
N_2	731.2	2.00	1612	2.00	--	--	--	--	--	--	--	--
H_2S	--	--	--	--	--	--	--	--	--	--	--	--
COS	--	--	--	--	--	--	--	--	--	--	--	--
H_2O	--	--	--	--	14908.7	100.00	32868	100.00	14197.4	100.00	31300	100.00
O_2	40930.4	98.00	90236	98.00	--	--	--	--	--	--	--	--
SO_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL												
	Kg/hr	41661.6	lb/hr	91848	Kg/hr	14908.7	lb/hr	32868	Kg/hr	14197.4	lb/hr	31300
	NM^3/hr	30885	SCFM	18176	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--
	Kl/hr	--	GPM	--	Kl/hr	14.93	GPM	66	Kl/hr	--	GPM	--
Temperature	C	110	F	230	C	110	F	230	C	135	F	275
Pressure	atm	2	PSIG	15	atm	--	PSIG	--	atm	3	PSIG	30

TABLE B-1.(Continued)

Stream No.	7				8				9			
Description	Spray Cooling Water				BFW to WH Boiler				Steam from WH Boiler			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%	Kg.hr	Mol.%	lb/hr	Mol.%
C	--	--	--	--	--	--	--	--	--	--	--	--
H	--	--	--	--	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--	--	--	--	--
S	--	--	--	--	--	--	--	--	--	--	--	--
O	--	--	--	--	--	--	--	--	--	--	--	--
Cl	--	--	--	--	--	--	--	--	--	--	--	--
Ash	--	--	--	--	--	--	--	--	--	--	--	--
CO	--	--	--	--	--	--	--	--	--	--	--	--
CO_2	--	--	--	--	--	--	--	--	--	--	--	--
H_2	--	--	--	--	--	--	--	--	--	--	--	--
N_2	--	--	--	--	--	--	--	--	--	--	--	--
H_2S	--	--	--	--	--	--	--	--	--	--	--	--
COS	--	--	--	--	--	--	--	--	--	--	--	--
H_2O	16456.3	100.0	36280	100.00	79855.8	100.00	176052	100.00	76047.5	100.00	167668	100.00
O_2	--	--	--	--	--	--	--	--	--	--	--	--
SO_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL												
	Kg/hr	16456.3	lb/hr	36280	Kg/hr	79855.8	lb/hr	176052	Kg/hr	76052.9	lb/hr	167668
	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--
	Kl/hr	16.48	GPM	72	Kl/hr	79.96	GPM	352	Kl/hr	--	GPM	--
Temperature	C	29	F	85	C	110	F	230	C	262	F	503
Pressure	atm	--	PSIG	--	atm	--	PSIG	--	atm	47.6	PSIG	685

TABLE B-1 (Continued)

Stream No.	10				11				12			
Description	Raw Gas to Scrubber				Gas to Gas Cooler				Scrubber Feed Water			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%
C	1745.4	--	3848	--	--	--	--	--	--	--	--	--
H	--	--	--	--	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--	--	--	--	--
S	--	--	--	--	--	--	--	--	--	--	--	--
O	--	--	--	--	--	--	--	--	--	--	--	--
Cl	--	--	--	--	--	--	--	--	--	--	--	--
Ash	3539.8	--	7804	--	--	--	--	--	--	--	--	--
CO	65712	41.44	144872	41.44	65712.8	40.18	144872	40.19	--	--	--	--
CO_2	18301.5	7.35	40348	7.35	18301.5	7.12	40348	7.12	--	--	--	--
H_2	2643.5	23.17	5828	23.17	2643.5	22.47	5828	22.47	--	--	--	--
N_2	1083.2	0.68	2388	0.68	1083.2	0.65	2388	0.65	--	--	--	--
H_2S	845.5	0.46	1864	0.46	845.5	0.44	1864	0.44	--	--	--	--
COS	108.9	0.034	240	0.034	108.9	0.032	240	0.032	--	--	--	--
H_2O	27397	26.86	60400	26.86	30622	29.1	67512	29.11	126579.5	100.00	67512	29.11
O_2	--	--	--	--	--	--	--	--	--	--	--	--
SO_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL	Kg/hr	121377.6	lb/hr	263052	Kg/hr	119318.3	lb/hr	263052	Kg/hr	126579.5	lb/hr	279060
	NM^3/hr	133951	SCFM	78832	NM^3/hr	138140	SCFM	81297	NM^3/hr	--	SCFM	--
	Kl/hr	--	GPM	--	Kl/hr	--	GPM	--	Kl/hr	126.7	GPM	558
Temperature	C	177	F	350	C	77	F	170	C	29	F	85
Pressure	atm	1.47	PSIG	6.9	atm	1.40	PSIG	5.9	atm	--	PSIG	--

TABLE B-1 (Continued)

Stream No.	13				14				15			
Description	Scrubbing Water Return				Gas to H_2S Removal				CW to Gas Cooler			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%
C	1745.4	1.36	3848	1.36	--	--	--	--	--	--	--	--
H	--	--	--	--	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--	--	--	--	--
S	--	--	--	--	--	--	--	--	--	--	--	--
O	--	--	--	--	--	--	--	--	--	--	--	--
Cl	--	--	--	--	--	--	--	--	--	--	--	--
Ash	3539.8	2.75	7804	2.75	--	--	--	--	--	--	--	--
CO	--	--	--	--	65712.3	54.37	144872	54.37	--	--	--	--
CO_2	--	--	--	--	18301.5	9.64	40348	9.64	--	--	--	--
H_2	--	--	--	--	2643.5	30.39	5828	30.39	--	--	--	--
N_2	--	--	--	--	1083.2	0.88	2388	0.88	--	--	--	--
H_2S	--	--	--	--	845.5	0.58	1864	0.58	--	--	--	--
COS	--	--	--	--	108.9	0.042	240	0.042	--	--	--	--
H_2O	123353.5	95.89	271948	95.89	3173.3	4.08	6996	4.08	921231.6	100.00	2030968	100.00
O_2	--	--	--	--	--	--	--	--	--	--	--	--
SO_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL	Kg/hr	128638.8	lb/hr	283600	Kg/hr	91868.7	lb/hr	202536	Kg/hr	9121231.6	lb/hr	2030968
	NM^3/hr	--	SCFM	--	NM^3/hr	102095	SCFM	60084	NM^3/hr	--	SCFM	--
	Kl/hr	123.5	GPM	544	Kl/hr	--	GPM	--	Kl/hr	922.4	GPM	4062
Temperature	C	49	F	120	C	35	F	95	C	29	F	85
Pressure	atm	--	PSIG	--	atm	1.34	PSIG	5	atm	--	PSIG	--

TABLE B-1 (Continued)

Stream No.	16				17				18			
Description	Gas Cooler Effluent Water				Claus Plant Feed				Clean Gas			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%
C	--	--	--	--	--	--	--	--	--	--	--	--
H	--	--	--	--	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--	--	--	--	--
S	--	--	--	--	--	--	--	--	--	--	--	--
O	--	--	--	--	--	--	--	--	--	--	--	--
Cl	--	--	--	--	--	--	--	--	--	--	--	--
Ash	--	--	--	--	--	--	--	--	--	--	--	--
CO	--	--	--	--	--	--	--	--	65712.8	56.10	144872	56.10
CO_2	--	--	--	--	2075.6	65.00	4576	65.00	16303.9	8.86	35944	8.86
H_2	--	--	--	--	--	--	--	--	2643.5	31.35	5828	31.35
N_2	--	--	--	--	--	--	--	--	1083.2	0.92	2388	0.92
H_2S	--	--	--	--	865.6	35.00	1908	35.00	41.7	0.03	92	0.03
COS	--	--	--	--	--	--	--	--	--	--	--	--
H_2O	948681.2	100.00	2091484	100.00	--	--	--	--	2062.9	2.74	4548	2.74
O_2	--	--	--	--	--	--	--	--	--	--	--	--
SO_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL												
	Kg/hr	948681.2	lb/hr	2091484	Kg/hr	2941.1	lb/hr	6484	Kg/hr	87848	lb/hr	193672
	NM^3/hr	--	SCFM	--	NM^3/hr	1717	SCFM	1010	NM^3/hr	98961	SCFM	58240
	Kl/hr	949.9	GPM	4183	Kl/hr	--	GPM	--	Kl/hr	--	GPM	--
Temperature	C	49	F	120	C	35	F	95	C	27	F	80
Pressure	atm	--	PSIG	--	atm	--	PSIG	--	atm	1.24	PSIG	3.5

TABLE B-1 (Continued)

Stream No.	19				20				21			
Description	Sulfur By-Product				Clean Gas to Air Heater				Clean Gas to Compressor			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%
C	--	--	--	--	--	--	--	--	--	--	--	--
H	--	--	--	--	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--	--	--	--	--
S	772.9	100.00	772.9	100.00	--	--	--	--	--	--	--	--
O	--	--	--	--	--	--	--	--	--	--	--	--
Cl	--	--	--	--	--	--	--	--	--	--	--	--
Ash	--	--	--	--	--	--	--	--	--	--	--	--
CO	--	--	--	--	3655.9	56.10	8060	56.10	62056.9	56.10	136812	56.10
CO_2	--	--	--	--	907.2	8.86	2000	8.86	15396.7	8.86	33944	8.86
H_2	--	--	--	--	146.9	31.35	324	31.35	2496.6	31.35	5504	31.35
N_2	--	--	--	--	59.9	0.92	132	0.92	1023.3	0.92	2256	0.92
H_2S	--	--	--	--	1.8	0.03	4	0.03	39.9	0.03	88	0.03
COS	--	--	--	--	--	--	--	--	--	--	--	--
H_2O	--	--	--	--	114.3	2.74	252	2.74	1948.6	2.74	4296	2.74
O_2	--	--	--	--	--	--	--	--	--	--	--	--
SO_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL	Kg/hr	772.9	lb/hr	1704	Kg/hr	4886	lb/hr	10772	Kg/hr	82962	lb/hr	182900
	NM^3/hr	--	SCFM	--	NM^3/hr	5505	SCFM	3240	NM^3/hr	93456	SCFM	5500
	Kl/hr	--	GPM	--	Kl/hr	--	GPM	--	Kl/hr	--	GPM	--
Temperature	C	--	F	--	C	27	F	80	C	27	F	80
Pressure	atm	--	PSIG	--	atm	1.24	PSIG	35	atm	1.24	PSIG	35

TABLE B-1 (Continued)

Stream No.	22				23				24			
Description	Hot Air				Air to Air Heater				Effluent from Coal Preparation			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%
C	--	--	--	--	--	--	--	--	--	--	--	--
H	--	--	--	--	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--	--	--	--	--
S	--	--	--	--	--	--	--	--	--	--	--	--
O	--	--	--	--	--	--	--	--	--	--	--	--
Cl	--	--	--	--	--	--	--	--	--	--	--	--
Ash	--	--	--	--	--	--	--	--	--	--	--	--
CO	--	--	--	--	--	--	--	--	--	--	--	--
CO_2	6651.5	2.12	14664	2.12	--	--	--	--	6651.5	1.88	14664	1.88
H_2	--	--	--	--	--	--	--	--	--	--	--	--
N_2	154620.6	77.58	340880	77.58	154558.9	79.00	340744	79.00	154670.6	68.57	340880	68.57
H_2S	--	--	--	--	--	--	--	--	--	--	--	--
COS	--	--	--	--	--	--	--	--	--	--	--	--
H_2O	1429.7	1.12	3152	1.12	--	--	--	--	18274.3	12.59	40288	12.59
O_2	43693.7	19.18	96328	19.18	46952.2	21.00	103512	21.00	43693.6	16.96	96328	16.96
SO_2	0.9	.0009	2	.0009	--	--	--	--	--	--	--	--
TOTAL	Kg/hr	206395.4	lb/hr	455024	Kg/hr	201511.1	lb/hr	444256	Kg/hr	223040	lb/hr	492160
	NM^3/hr	168438	SCFM	99128	NM^3/hr	165339	SCFM	97304	NM^3/hr	190555	SCFM	112144
	Kl/hr	--	GPM	--	Kl/hr	--	GPM	--	Kl/hr	--	GPM	--
Temperature	C	260	F	500	C	--	F	--	C	74	F	165
Pressure	atm	--	PSIG	--	atm	--	PSIG	--	atm	--	PSIG	--

TABLE B-1 (Continued)

Stream No.	25				26				27			
Description	Blowdown Water				Blowdown Water				Slug to Filter			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%
C	--	--	--	--	--	--	--	--	--	--	--	--
H	--	--	--	--	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--	--	--	--	--
S	--	--	--	--	--	--	--	--	--	--	--	--
O	--	--	--	--	--	--	--	--	--	--	--	--
Cl	--	--	--	--	--	--	--	--	--	--	--	--
Ash	--	--	--	--	--	--	--	--	3539.8	50.00	7804	50.00
CO	--	--	--	--	--	--	--	--	--	--	--	--
CO_2	--	--	--	--	--	--	--	--	--	--	--	--
H_2	--	--	--	--	--	--	--	--	--	--	--	--
N_2	--	--	--	--	--	--	--	--	--	--	--	--
H_2S	--	--	--	--	--	--	--	--	--	--	--	--
COS	--	--	--	--	--	--	--	--	--	--	--	--
H_2O	709.4	100.00	1564	100.00	3802.9	100.00	8384	100.00	3539.8	50.00	7804	50.00
O_2	--	--	--	--	--	--	--	--	--	--	--	--
SO_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL												
	Kg/hr	709.4	lb/hr	1564	Kg/hr	3802.9	lb/hr	8384	Kg/hr	7079.6	lb/hr	15608
	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--
	Kl/hr	0.71	GPM	3	Kl/hr	3.81	GPM	17	Kl/hr	--	GPM	--
Temperature	C	--	F	--	C	--	F	--	C	49	F	120
Pressure	atm	--	PSIG	--	atm	--	PSIG	--	atm	--	PSIG	--

TABLE B-1 (Continued)

Stream No.	28				29				30			
Description	Cooling Water to Quench Tank				Water to Clarifier				Slurry to Filter			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	wt%
C	--	--	--	--	--	--	--	--	--	--	--	--
H	--	--	--	--	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--	--	--	--	--
S	--	--	--	--	--	--	--	--	--	--	--	--
O	--	--	--	--	--	--	--	--	--	--	--	--
Cl	--	--	--	--	--	--	--	--	--	--	--	--
Ash	--	--	--	--	--	--	--	--	3539.8	20.09	7804	20.09
CO	--	--	--	--	--	--	--	--	--	--	--	--
CO_2	--	--	--	--	--	--	--	--	--	--	--	--
H_2	--	--	--	--	--	--	--	--	--	--	--	--
N_2	--	--	--	--	--	--	--	--	--	--	--	--
H_2S	--	--	--	--	--	--	--	--	--	--	--	--
COS	--	--	--	--	--	--	--	--	--	--	--	--
H_2O	77571.5	100.00	171016	100.00	74031.7	100.00	163212	100.00	12332.3	70.00	27183	70.00
O_2	--	--	--	--	--	--	--	--	--	--	--	--
SO_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL												
	Kg/hr	77571.5	lb/hr	171016	Kg/hr	74031.7	lb/hr	163212	Kg/hr	17617.5	lb/hr	38840
	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--
	Kl/hr	77.67	GPM	342	Kl/hr	74.12	GPM	326	Kl/hr	12.4	GPM	54.4
Temperature	C	29	F	85	C	49	F	120	C	29	F	85
Pressure	atm	--	PSIG	--	atm	--	PSIG	--	atm	--	PSIG	--

TABLE B-1 (Continued)

Stream No.	31				32				33			
Description	Slag				Water from Filter				Water to Cooling Tower			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	wt%	lb/hr	wt%	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%
C	1745.5	17.80	3848	17.80	--	--	--	--	--	--	--	--
H	--	--	--	--	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--	--	--	--	--
S	--	--	--	--	--	--	--	--	--	--	--	--
O	--	--	--	--	--	--	--	--	--	--	--	--
Cl	--	--	--	--	--	--	--	--	--	--	--	--
Ash	7079.6	72.20	15608	72.20	--	--	--	--	--	--	--	--
CO	--	--	--	--	--	--	--	--	--	--	--	--
CO_2	--	--	--	--	--	--	--	--	--	--	--	--
H_2	--	--	--	--	--	--	--	--	--	--	--	--
N_2	--	--	--	--	--	--	--	--	--	--	--	--
H_2S	--	--	--	--	--	--	--	--	--	--	--	--
COS	--	--	--	--	--	--	--	--	--	--	--	--
H_2O	979.8	10.00	2160	10.00	14892.3	100.00	32832	100.00	50122	100.00	110500	100.00
O_2	--	--	--	--	--	--	--	--	--	--	--	--
SO_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL												
	Kg/hr	9804.8	lb/hr	21616	Kg/hr	14892.3	lb/hr	32832	Kg/hr	50122	lb/hr	110500
	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--
	Kl/hr	--	GPM	--	Kl/hr	14.9	GPM	66	Kl/hr	50.2	GPM	221
Temperature	C	--	F	--	C	29	F	85	C	29	F	85
Pressure	atm	--	PSIG	--	atm	--	PSIG	--	atm	--	PSIG	--

TABLE B-1 (Continued)

Stream No.	34				35			
Description	Make-Up Water				Clean Fuel Gas			
	Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%
C	--	--	--	--	--	--	--	--
H	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--
S	--	--	--	--	--	--	--	--
O	--	--	--	--	--	--	--	--
Cl	--	--	--	--	--	--	--	--
Ash	--	--	--	--	--	--	--	--
CO	--	--	--	--	62056.9	56.88	136812	56.88
CO_2	--	--	--	--	15396.7	8.98	33944	8.98
H_2	--	--	--	--	2496.6	31.79	5504	31.79
N_2	--	--	--	--	1023.3	0.94	2256	0.94
H_2S	--	--	--	--	39.9	0.03	88	0.03
COS	--	--	--	--	--	--	--	--
H_2O	31751.5	100.00	70000	100.00	968.9	1.38	2136	1.38
O_2	--	--	--	--	--	--	--	--
SO_2	--	--	--	--	--	--	--	--
TOTAL								
	Kg/hr	31751.5	lb/hr	70000	Kg/hr	81982.3	lb/hr	180740
	NM^3/hr	--	SCFM	--	NM^3/hr	92171.4	SCFM	54244
	Kl/hr	31.8	GPM	85	Kl/hr	--	GPM	--
Temperature	C	--	F	--	C	16	F	60
Pressure	atm	--	PSIG	--	atm	6.44	PSIG	80

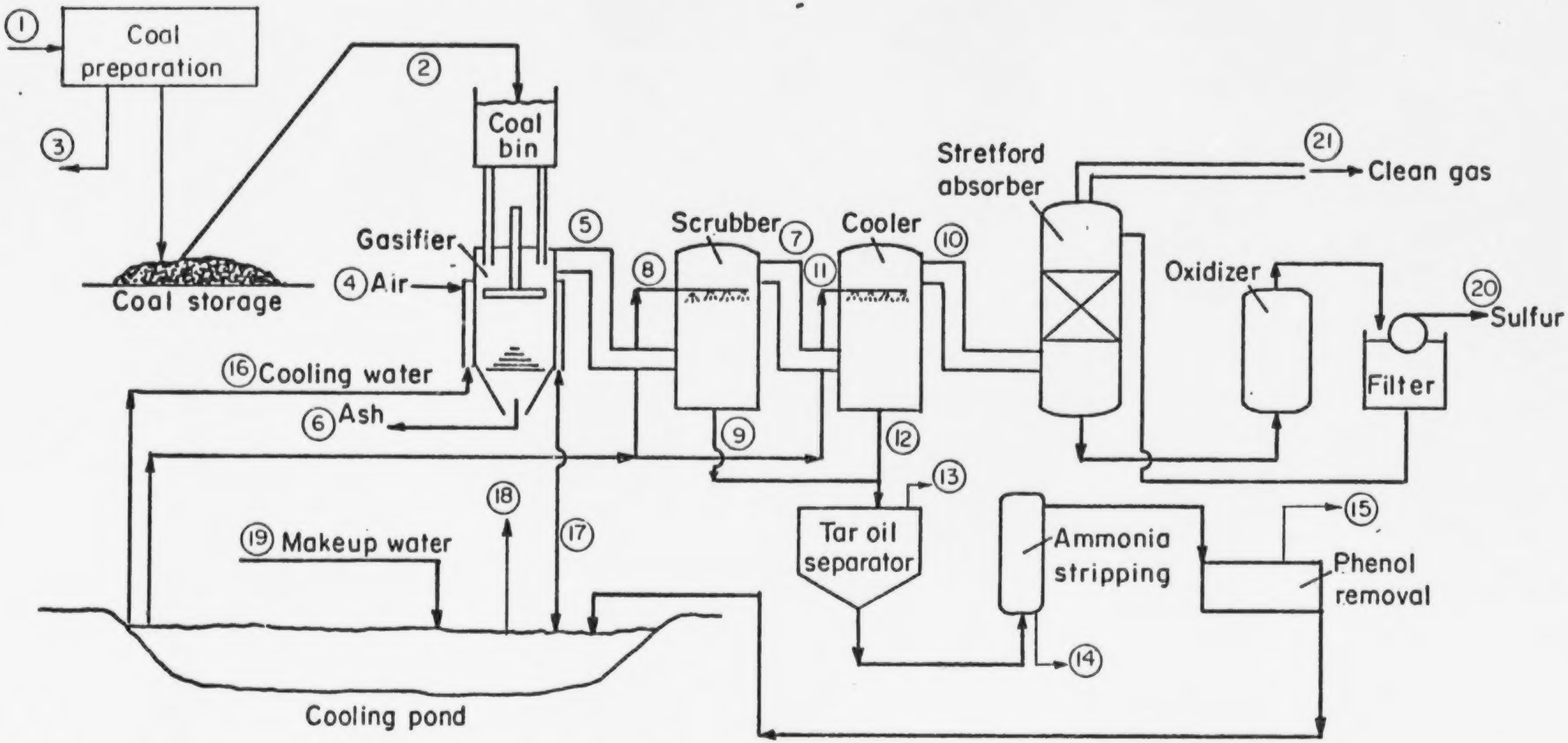
① Coal preparation
③
② Coal bin
Coal storage
Gasifier
④ Air
⑤
⑯ Cooling water
⑥ Ash
⑧
Scrubber ⑦
⑨
⑪
Cooler ⑩
⑫
⑬
Tar oil separator
Stretford absorber
㉑ Clean gas
Oxidizer
Filter
⑳ Sulfur
Ammonia stripping
⑭
⑮
Phenol removal
⑱
⑰
⑲ Makeup water
Cooling pond

FIGURE B-2. WELLMAN-GALUSHA/STRETFORD GASIFICATION PLANT MATERIAL BALANCE FOR MODEL REFINERY PLANT

TABLE B-2. WELLMAN-GALUSHA/STRETFORD GASIFICATION PLANT MATERIAL BALANCE FOR MODEL REFINERY PLANT
(Metric and English Units)

Stream No.	1				2				3			
Description	Coal as Received				Coal Fines				Coal Feed to Gasifier			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	wt%	lb/hr	wt%	Kg/hr	Wt%	lb/hr	wt%	Kg/hr	wt%	lb/hr	wt%
C	7711.1	68.0	17000	68.0	1233.8	68.0	2720	68.0	6477.3	68.0	14280	68.0
H	544.3	4.8	1200	4.8	87.1	4.8	192	4.8	457.2	4.8	1008	4.8
N	238.1	2.1	525	2.1	38.1	2.1	84	2.1	200.0	2.1	441	2.1
S	442.3	3.9	975	3.9	70.8	3.9	975	3.9	371.5	3.9	156	3.9
O	771.1	6.8	1700	6.8	123.4	6.8	272	6.8	647.7	6.8	1428	6.8
Ash	952.5	8.4	2100	8.4	152.4	8.4	336	8.4	800.1	8.4	1764	8.4
CO	--	--	--	--	--	--	--	--	--	--	--	--
CO2	--	--	--	--	--	--	--	--	--	--	--	--
H_2	--	--	--	--	--	--	--	--	--	--	--	--
CH_4	--	--	--	--	--	--	--	--	--	--	--	--
N_2	--	--	--	--	--	--	--	--	--	--	--	--
NH_3	--	--	--	--	--	--	--	--	--	--	--	--
H_2S	--	--	--	--	--	--	--	--	--	--	--	--
Phenols	--	--	--	--	--	--	--	--	--	--	--	--
Tar	--	--	--	--	--	--	--	--	--	--	--	--
H_2O	680.4	6.0	1500	6.0	108.9	6.0	240	6.0	571.5	6.0	1260	6.0
O_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL	Kg/hr	11339.8	lb/hr	25000	Kg/hr	1814.4	lb/hr	4000	Kg/hr	9575.4	lb/hr	21000
	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--
	Kl/hr	--	GPM	--	Kl/hr	--	GPM	--	Kl/hr	--	GPM	--
Temperature	C	25	F	77	C	25	F	77	C	25	F	77
Pressure	atm	--	PSIG	--	atm	--	PSIG	--	atm	--	PSIG	--

TABLE B-2 (Continued)

Stream No.	4				5				6			
Description	Air and Steam				Hot Raw Gas				Ash			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	wt%	lb/hr	wt%
C	--	--	--	--	9.1	--	20	--	81.7	10.18	180	10.18
H	--	--	--	--	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--	--	--	--	--
S	--	--	--	--	--	--	--	--	--	--	--	--
O	--	--	--	--	--	--	--	--	--	--	--	--
Ash	--	--	--	--	79.8	--	176	--	720.3	89.82	1588	89.82
CO	--	--	--	--	11383.8	24.86	25097	24.86	--	--	--	--
CO_2	--	--	--	--	2126.0	2.95	4687	2.95	--	--	--	--
H_2	--	--	--	--	421.9	12.77	930	12.77	--	--	--	--
CH_4	--	--	--	--	615.5	2.35	1357	2.35	--	--	--	--
N_2	19186.0	59.01	42298	59.01	19300.8	42.17	42551	42.17	--	--	--	--
NH_3	--	--	--	--	99.3	0.36	219	0.36	--	--	--	--
H_2S	--	--	--	--	389.2	0.70	858	0.70	--	--	--	--
Phenols	--	--	--	--	54.4	--	120	--	--	--	--	--
Tar	--	--	--	--	475.4	--	1048	--	--	--	--	--
H_2O	5292.5	25.30	11668	25.30	4075.5	13.84	8985	13.84	--	--	--	--
O_2	5828.7	15.69	12850	15.69	--	--	--	--	--	--	--	--
TOTAL	Kg/hr	30307.2	lb/hr	66816	Kg/hr	39030.7	lb/hr	86048	Kg/hr	802.0	lb/hr	1768
	NM^3/hr	27474	SCFM	16169	NM^3/hr	38681	SCFM	22764	NM^3/hr	--	SCFM	--
	Kl/hr	--	GPM	--	Kl/hr	--	GPM	--	Kl/hr	--	GPM	--
Temperature	C	66	F	150	C	360	F	680	C	--	F	--
Pressure	atm	--	PSIG	--	atm	--	PSIG	--	atm	--	PSIG	--

TABLE B-2 (Continued)

Stream No.	7				8				9			
Description	Gas to Gas Cooler				Scrubbing Water				Scrubbing Water Return			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	wt%	lb/hr	wt%
C	--	--	--	--	--	--	--	--	9.1	0.01	20	0.01
H	--	--	--	--	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--	--	--	--	--
S	--	--	--	--	--	--	--	--	--	--	--	--
O	--	--	--	--	--	--	--	--	--	--	--	--
Ash	--	--	--	--	--	--	--	--	79.8	0.12	176	0.12
CO	11383.8	23.90	25097	23.94	--	--	--	--	--	--	--	--
CO_2	2126.0	2.84	4687	2.84	--	--	--	--	--	--	--	--
H_2	421.9	12.30	930	12.30	--	--	--	--	--	--	--	--
CH_4	615.5	2.26	1357	2.26	--	--	--	--	--	--	--	--
N_2	19300.8	40.60	42551	40.60	--	--	--	--	--	--	--	--
NH_3	--	--	--	--	--	--	--	--	99.3	0.14	219	0.14
H_2S	389.2	0.67	858	0.67	--	--	--	--	--	--	--	--
Phenols	--	--	--	--	--	--	--	--	54.4	0.08	120	0.08
Tar	--	--	--	--	--	--	--	--	475.4	0.69	1048	0.69
H_2O	5320.6	17.39	11730	17.39	69313.5	100.00	152810	100.00	68068.4	98.96	150065	98.96
O_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL	Kg/hr	39557.8	lb/hr	87210	Kg/hr	69313.5	lb/hr	152810	Kg/hr	68786.4	lb/hr	151648
	NM^3/hr	40178	SCFM	23645	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--
	Kl/hr	--	GPM	--	Kl/hr	69.4	GPM	68.8	Kl/hr	77	GPM	127
Temperature	C	57	F	135	C	25	F	77	C	53	F	127
Pressure	atm	--	PSIG	--	atm	--	PSIG	--	atm	--	PSIG	--

TABLE B-2 (Continued)

Stream No.	10				11				12			
Description	Gas to Sulfur Removal				Cooling Water				Cooling Water Return			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%
C	--	--	--	--	--	--	--	--	--	--	--	--
H	--	--	--	--	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--	--	--	--	--
S	--	--	--	--	--	--	--	--	--	--	--	--
O	--	--	--	--	--	--	--	--	--	--	--	--
Ash	--	--	--	--	--	--	--	--	--	--	--	--
CO	11383.8	27.35	25097	27.35	--	--	--	--	--	--	--	--
CO_2	2126.0	3.25	4687	3.25	--	--	--	--	--	--	--	--
H_2	421.9	14.06	930	14.06	--	--	--	--	--	--	--	--
CH_4	615.5	2.58	1357	2.58	--	--	--	--	--	--	--	--
N_2	19300.8	46.39	42551	46.39	--	--	--	--	--	--	--	--
NH_3	--	--	--	--	--	--	--	--	--	--	--	--
H_2S	389.2	0.77	858	0.77	--	--	--	--	--	--	--	--
Phenols	--	--	--	--	--	--	--	--	--	--	--	--
Tar	--	--	--	--	--	--	--	--	--	--	--	--
H_2O	1499.6	5.60	3306	5.60	125928.6	100.00	277625	100.00	129749.6	100.00	286049	100.00
O_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL	Kg/hr NM^3/hr Kl/hr	35736.7 35160 --	lb/hr SCFM GPM	78786 20692 --	Kg/hr NM^3/hr Kl/hr	125928.6 -- 126.1	lb/hr SCFM GPM	277625 -- 555	Kg/hr NM^3/hr Kl/hr	129749.6 -- 129.9	lb/hr SCFM GPM	286049 -- 572
Temperature	C	35	F	95	C	25	F	77	C	53	F	127
Pressure	atm	--	PSIG	--	atm	--	PSIG	--	atm	--	PSIG	--

TABLE B-2 (Continued)

Stream No.	13				14				15			
Description	Tar				Ammonia Solution				Phenols			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	wt%	lb/hr	wt%	Kg/hr	wt%	lb/hr	wt%	Kg/hr	Mol.%	lb/hr	Mol.%
C	422.8	80.83	932	80.83	--	--	--	--	--	--	--	--
H	36.7	7.03	81	7.03	--	--	--	--	--	--	--	--
N	3.6	0.69	8	0.69	--	--	--	--	--	--	--	--
S	5.5	1.04	12	1.04	--	--	--	--	--	--	--	--
O	6.8	1.30	15	1.30	--	--	--	--	--	--	--	--
Ash	--	--	--	--	--	--	--	--	--	--	--	--
CO	--	--	--	--	--	--	--	--	--	--	--	--
CO_2	--	--	--	--	--	--	--	--	--	--	--	--
H_2	--	--	--	--	--	--	--	--	--	--	--	--
CH_4	--	--	--	--	--	--	--	--	--	--	--	--
N_2	--	--	--	--	--	--	--	--	--	--	--	--
NH_3	--	--	--	--	99.3	20.00	219	20.00	--	--	--	--
H_2S	--	--	--	--	--	--	--	--	--	--	--	--
Phenols	--	--	--	--	--	--	--	--	54.4	--	120	--
Tar	--	--	--	--	--	--	--	--	--	--	--	--
H_2O	47.6	9.11	105	9.11	397.4	80.00	876	80.00	--	--	--	--
O_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL	Kg/hr	523	lb/hr	1153	Kg/hr	496.7	lb/hr	1095	Kg/hr	54.4	lb/hr	120
	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--
	Kl/hr	--	GPM	--	Kl/hr	0.42	GPM	2	Kl/hr	--	GPM	--
Temperature	C	25	F	77	C	25	F	77	C	--	F	--
Pressure	atm	--	PSIG	--	atm	--	PSIG	--	atm	--	PSIG	--

TABLE B-2 (Continued)

Stream No.	16				17				18			
Description	Water to Gasifier Jacket				Water from Gasifier Jacket				Water Loss			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%
C	--	--	--	--	--	--	--	--	--	--	--	--
H	--	--	--	--	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--	--	--	--	--
S	--	--	--	--	--	--	--	--	--	--	--	--
O	--	--	--	--	--	--	--	--	--	--	--	--
Ash	--	--	--	--	--	--	--	--	--	--	--	--
CO	--	--	--	--	--	--	--	--	--	--	--	--
CO_2	--	--	--	--	--	--	--	--	--	--	--	--
H_2	--	--	--	--	--	--	--	--	--	--	--	--
CH_4	--	--	--	--	--	--	--	--	--	--	--	--
N_2	--	--	--	--	--	--	--	--	--	--	--	--
NH_3	--	--	--	--	--	--	--	--	--	--	--	--
H_2S	--	--	--	--	--	--	--	--	--	--	--	--
Phenols	--	--	--	--	--	--	--	--	--	--	--	--
Tar	--	--	--	--	--	--	--	--	--	--	--	--
H_2O	64259.5	100.00	141668	100.00	58976	100.00	130000	100.00	13519.3	100.00	29805	100.00
O_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL	Kg/hr NM^3/hr Kl/hr	64259.5 -- 64.3	lb/hr SCFM GPM	141668 -- 283	Kg/hr NM^3/hr Kl/hr	58967 -- 59.0	lb/hr SCFM GPM	130000 -- 260	Kg/hr NM^3/hr Kl/hr	13519.3 -- 13.5	lb/hr SCFM GPM	29805 -- 260
Temperature	C	25	F	77	C	66	F	150	C	--	F	--
Pressure	atm	--	PSIG	--	atm	--	PSIG	--	atm	--	PSIG	--

TABLE B-2 (Continued)

Stream No.	19				20				21			
Description	Make-Up Water				Sulfur				Clean Gas			
	Metric Unit		English Unit		Metric Unit		English Unit		Metric Unit		English Unit	
Composition	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%	Kg/hr	Mol.%	lb/hr	Mol.%
C	--	--	--	--	--	--	--	--	--	--	--	--
H	--	--	--	--	--	--	--	--	--	--	--	--
N	--	--	--	--	--	--	--	--	--	--	--	--
S	--	--	--	--	352.4	100.00	777	100.00	--	--	--	--
O	--	--	--	--	--	--	--	--	--	--	--	--
Ash	--	--	--	--	--	--	--	--	--	--	--	--
CO	--	--	--	--	--	--	--	--	11383.8	28.70	25097	28.70
CO_2	--	--	--	--	--	--	--	--	2126.0	3.41	4687	3.41
H_2	--	--	--	--	--	--	--	--	421.9	14.74	930	14.74
CH_4	--	--	--	--	--	--	--	--	615.5	2.71	1357	2.71
N_2	--	--	--	--	--	--	--	--	19300.8	48.67	42551	48.67
NH_3	--	--	--	--	--	--	--	--	--	--	--	--
H_2S	--	--	--	--	--	--	--	--	14.5	0.03	32	0.03
Phenols	--	--	--	--	--	--	--	--	--	--	--	--
Tar	--	--	--	--	--	--	--	--	--	--	--	--
H_2O	16680.9	100.00	36775	100.00	--	--	--	--	444.1	1.74	979	1.74
O_2	--	--	--	--	--	--	--	--	--	--	--	--
TOTAL	Kg/hr	16680.9	lb/hr	36775	Kg/hr	352.4	lb/hr	777	Kg/hr	34306.6	lb/hr	75633
	NM^3/hr	--	SCFM	--	NM^3/hr	--	SCFM	--	NM^3/hr	33513	SCFM	19723
	Kl/hr	16.7	GPM	73.5	Kl/hr	--	GPM	--	Kl/hr	--	GPM	--
Temperature	C	25	F	77	C	--	F	--	C	16	F	60
Pressure	atm	--	PSIG	--	atm	--	PSIG	--	atm	--	PSIG	--